August 2010

COMSOL Assisted Modeling of a Climbing Film Evaporator

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Project Number: WMC 2228

COMSOL Assisted Modeling of a Climbing Film Evaporator

A Major Qualifying Project Report

Submitted to the Faculty

of the

WORCESTER POLYTECHNIC INSTITUTE

in partial fulfillment of the requirements for the

Degree of Bachelor of Science

by

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Date: August 27, 2010

Approved:

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Professor William M. Clark, Project Advisor

Abstract

The aim of this project was to create a computer model of a climbing film evaporator using COMSOL Multiphysics. We used the climbing film evaporator lab designed for CHE 4402 to structure our lab experiment and to collect lab data about the change of composition in the evaporator feed solution. The data was then used to create a simulation model using COMSOL Multiphysics. COMSOL approximated the experimental data very well predicting a product concentration of 12.8 percent glycerol in water whereas in the experiment the measured concentration was 12 percent. The energy balance results did not match very closely with COMSOL reporting 5775 W of heat given by the steam whereas in the experiment the calculated heat given by the steam was 5909 W and COMSOL predicted a steam trap rate of 9.1 kg/h and the experimentally obtained data was 9.6 kg/h. We concluded that COMSOL can be an effective tool for simulating a climbing film evaporator given the correct heat transfer coefficients, heat flux expressions, boundary conditions, and concentrations and we developed recommendations, which we present regarding future modeling and experimentation.

Acknowledgements

The success of this project depended on the contributions of two individuals over the past eight months. I would like to take time to thank all of those who have helped and supported me in this process.

First, we want to thank my advisor, Professor William M. Clark for his guidance throughout the project. His supervision during the laboratory experiment and his inputs on the development of the COMSOL model were invaluable.

Secondly, I would like to thank my project partner, Miguel Herrera for his constant help and contributions to help complete this project.

Next, I would like to thank Mr. Jack Ferraro for setting up the experiment and always being available for assistance.

Contents

Table of Figures

Table of Tables

Introduction

With the development of technology powerful simulation softwares have become more readily available to the public. Computer simulation makes it easier for people to better understand complicated physical phenomena that occur in apparatuses used to design certain chemical engineering processes. This is possible because these simulations are able to provide visual representation of otherwise hard to picture concepts such as, concentration gradients, velocity profiles and temperature gradients. Although running these processes first hand in the laboratory is an excellent way to complement theoretical knowledge and understand the basic principles and theories behind these unit operations it can be very useful to have digital simulations that virtually model these processes and provide illustrations of basic chemical engineering principles virtually.

One such software is COMSOL Multiphysics. COMSOL is a finite element analysis and solver software package for coupled phenomena or Multiphysics. It is particularly good at modeling chemical engineering apparatus since it is specifically designed to easily combine transport phenomena, computational fluid dynamics and mass and energy transport to chemical reaction kinetics. COMSOL has the ability to solve multiple non linear PDE's simultaneously and the models can be generated and solved in one, two or even three dimensions [13]. COMSOL Multiphysics is a very helpful tool as the models are very interactive and user friendly and ideal tools to complement theoretical knowledge in classrooms, lab tutorials and study guides.

The objective of this project was to create a COMSOL model for a climbing film evaporator. In brief, a climbing film evaporator is a unit operation in which a solution is concentrated by removing a part of the solvent in the form of vapor [3]. The most commonly used solvent is water and the latent heat of evaporation is usually supplied by condensing steam. Heat from the steam is transmitted to the solution by conduction and convection through the glass wall of the evaporator. When the solvent starts boiling the bubbles inside the tube create an upward flow that causes the mixture to rise [6] and finally sloshes over to a container were the concentrated solution is collected. Similarly, the vaporized solvent is collected in a separate container after going through a condenser. Climbing film evaporators are widely used in the food and drink industry as means to concentrate fruit juices, coffee and tea. They are also used to recover expensive solvents from solutions that otherwise would be wasted.

This project builds on the work that was done in collaboration with Miguel Herrera, who is a co-author for this project. In order to create the model, first we performed an experiment using the climbing film evaporator located in the Unit Operations Laboratory following the designed experiment guidelines for course CHE 4402. In short, for the experiment we had a solution of 10 percent glycerol in water that was fed to the evaporator. Several runs were performed using different feed flow rates and steam pressures. We recorded the concentration of the product as well as the flow rate, the flow rate of the condensate, and the flow rate of the condenser solution. We performed mass balances and energy balances on the system to calculate heat transfer coefficients, heat lost to the environment and heat given by the steam. The results obtained from the experimental calculations were then used to create a COMSOL model of our climbing film evaporator. This project explains and illustrates the climbing film

evaporation process and strives to model the same behavior using COMSOL Multiphysics to aid future users in understanding the process.

Background and Theory

Background for Climbing Film Evaporator

Evaporation

The vaporization of a liquid for the purpose of concentrating a solution consisting of a solute and solvent of different volatilities (generally the solvent is more volatile than the solute) is a common unit operation and is performed in various industrial as well as domestic settings using many different methods. Evaporation is a common way to achieve this by vaporizing a part of the solvent in the form of vapor to obtain a concentrated product.

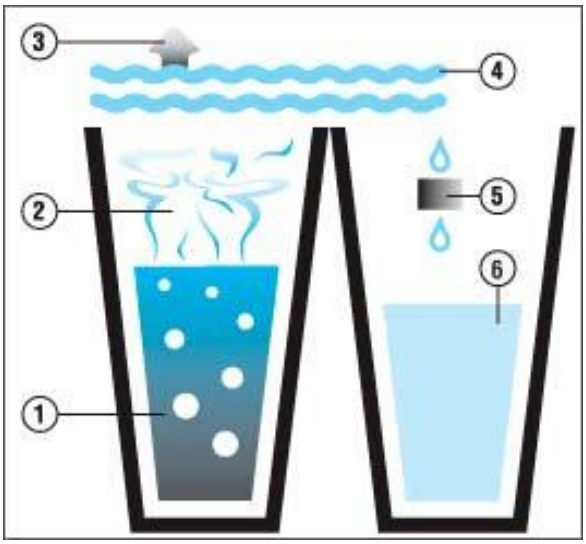

Figure 1: Example of concentrating a liquid by using evaporation as a unit operation [11].

In most industrial cases the solvent is water and the energy to evaporate this water is supplied by the latent heat of vaporization of condensing steam. This energy transfer takes place by indirect heat transfer through some conducting surfaces (such as metals or glass).

Evaporation differs from drying in the fact that in drying the residue is generally a solid whereas the residue in evaporation is a liquid, which is highly viscous in some cases. It differs from distillation in the fact that in distillation the vapor is generally a mixture of various components rather than a pure compound. In evaporation the vapor is usually a single component and even if it is, no attempt is made no attempt is made to separate the vapor into fractions.(This should not confused with the separation step in a Climbing film evaporator where the cyclone separator splits the feed into liquid and vapor components instead of fractionating the vapor). The conditions under which evaporators are used vary widely according to the application [9]. Sometimes the liquid to be evaporated is less viscous than the water whereas the other times it can be so viscous that it almost gel like or semi solid in nature. Similarly

different chemical and physical properties of the solvent affect the behavior it displays under the influence of heat or high temperatures. Some solids might leave a scale on the on the surface of the apparatus or might be thermally unstable or damageable under the influence of heat energy. These variations in the chemical compound behaviors and the applications of the process (industrial or domestic) have led to various types of designs for evaporators.

Evaporator types can be classified as [2]:

- Jacketed Vessels
- Coils
- Horizontal tube evaporators
- Short tube vertical evaporators
- Long tube vertical evaporators
    - Forced circulation
    - **Upward flow (climbing film)**
    - Downward flow (falling film)
- Forced Circulation Evaporators
- Flash Evaporators

Climbing Film Evaporator

A climbing film evaporator is a type of long tube vertical evaporator. A CFE is a shell and tube heat exchanger mounted to a vapor/liquid separator [Ref].

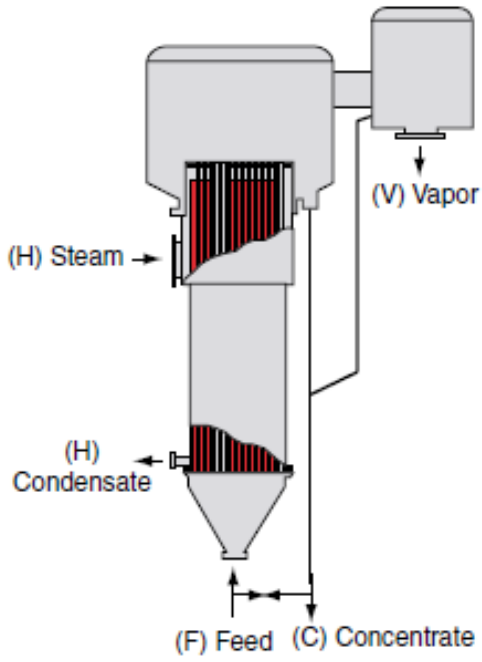

Figure 2: Image of a climbing film evaporator [6].

These evaporators are generally operated under vacuum in order to reduce the boiling point of the feed solution and increase the heat flux. The principle behind the CFE is that any kind of vapor (steam in our case) flowing at a higher velocity than the liquid (glycerol solution in our case) flows into the cavity between the two glass tubes of the evaporator causing the liquid to rise up the inner tube in a film[6]. The feed solution enters the bottom of the inner tube and flows upwards as a result of forced circulation due to a pump. In the lower section of the tube the feed solution is heated up to the boiling point of the solvent. At some height in the inner tube bubbles start to form indicating that the more volatile substance has attained its boiling point. This height is called the boiling height of the liquid. The ascending force of the water vapor produced during boiling causes the liquid and the vapor to rise upwards in parallel flow. At the same time the production of water vapor increases and the product starts to form a thin film on the walls of the inner tube of the evaporator and the liquid mixture begins to rise upwards. This co-current flow of the liquid and the vapor against gravity creates a high degree of turbulence in the liquid. This results higher linear velocity and rate of heat transfer and is beneficial during evaporation of highly viscous products or products that have a tendency to foul the surface of the evaporator [8]. In this boiling zone a mixture of vapor and liquid tend to rise quickly to the top of the tube and are discharged at high velocity from the top. They are sent into the cyclone separator which then separates them to be sent to the product line or condenser.

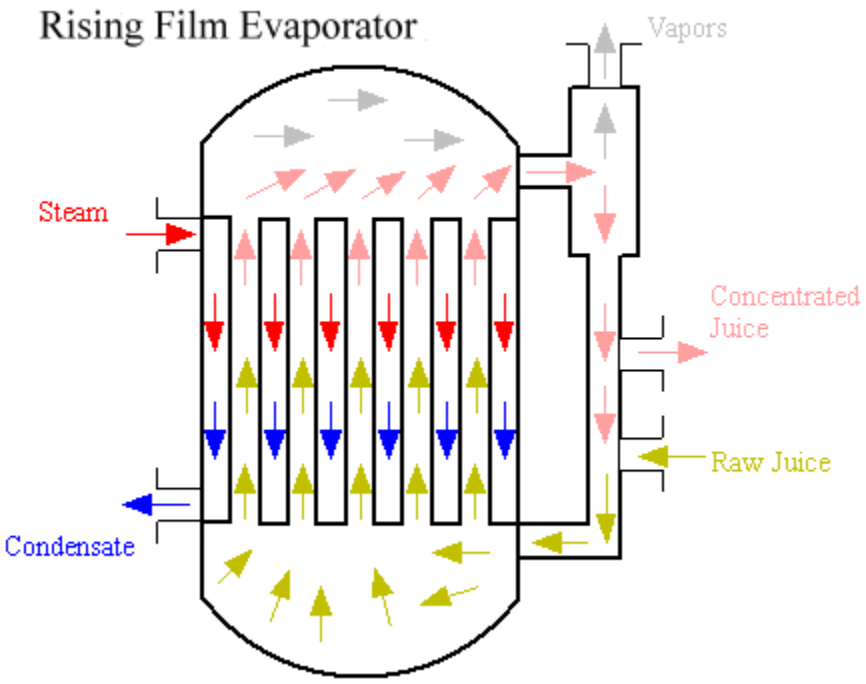

Figure 3: Schematic showing the flow of liquid and vapor in the CFE [10]

A lot of times this type of evaporator is used with product recirculation, where some of the concentrated product is recycled back into the feed solution (just like in cases of distillation) in order to concentrate the product further and produce sufficient liquid loading inside the heating tubes.

Advantages of using the climbing film evaporator include [6]:

- Reduced floor space requirements
- Higher heat transfer coefficient due to partial two-phase flow
- Ability to handle foamy liquids
- Low residence time which permits the use of CFE's for heat sensitive materials such as food products or thermally unstable chemicals.
- Another advantage of using the climbing film evaporator is the low cost of construction.

The disadvantages include:

- Higher pressure drop through the tube compared to other tubular evaporators
- High head- room requirements

Multitube CFE's are often used in the industry to concentrate solutions such as fruit juices that can be damaged by prolonged heat. Some of the most common uses of the CFE include concentration of cane sugar syrups, black liquor in paper plants, nitrates and electrolytic tinning liquors [2].

Each climbing film evaporator is set with certain major and minor equipment which are as follows [2]:

- A condenser
- Vacuum producing pump
- Condensate removing steam traps
- Process Pumps
- Process Piping
- Safety and Relief Equipment such as valves
- Thermal Insulation
- Process Vessels
- Electronic monitors and flow meters

Background for COMSOL

There are various unit operations which are used to design certain chemical engineering process whether they are in the industry or small scale laboratories. Although running these processes first hand in the laboratory is an excellent way to complement theoretical knowledge and understand the basic principles and theories behind these unit operations it can be very useful to have digital simulations that virtually model these processes and provide illustrations of basic chemical engineering principles virtually.

One such programming package used to simulate various chemical engineering processes is COMSOL Multiphysics. This is a finite element analysis and solver software package for coupled phenomena or Multiphysics [13]. There is a special chemical engineering module which is a great tool for process related modeling. It is specifically designed to easily combine transport phenomena, computational fluid dynamics and mass and energy transport to chemical reaction kinetics. COMSOL has the ability to solve multiple non linear PDE's simultaneously and the models can be generated in solved in one, two or even three dimensions [13]. COMSOL Multiphysics is a very helpful tool as the models are very interactive and

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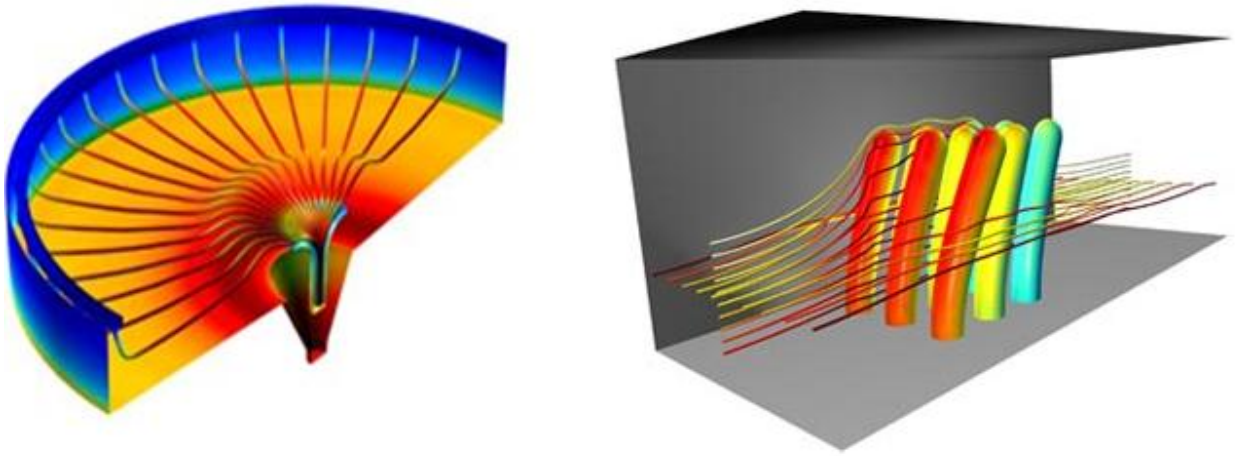


Figure 4: Different modules available in COMSOL [13]

The following are the basic steps to create a model in COMSOL:

Generate the geometry of the process that you want to simulate. This geometry can also be imported from other sources. Different geometries can be selected based on the dimensions of the process model. The geometry then requires to be meshed in order to create a grid of small, simple shaped data points that the program can solve. The size and type of mesh depend on the desired final process. After creating a meshed geometry the physics of the process being solved can be defined in the sub domain settings and then known values and constants can be entered to solve the model. Once the program has solved the model the post processing of the results enables the user to generate variation profiles, maps and plots of process variables. These can be extrapolated or interpolated in time or beyond parametric solutions.

Methodology

Part 1: Conducting the experiment on the Climbing Film Evaporator

Schematic of the Climbing Film Evaporator

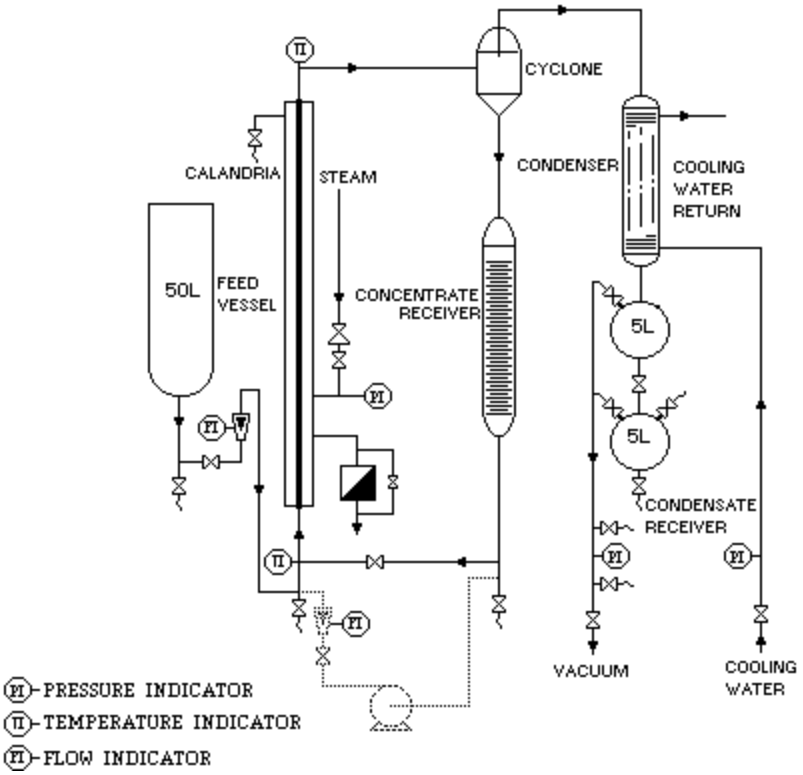


Figure 5: Schematic of the Climbing Film Evaporator [12]

Equipment Summary

The climbing film evaporator is a tower consisting of two concentric glass tubes. The dimensions of the CFE in Goddard Hall 116 were measure and the inner diameter was found to be 1 inch, the outer diameter was 2.5 inches and the length of the glass tubes was measured to be 9 ft. The evaporator is connected to a pump and rotameter which supply the feed solution to be concentrated into the inner tube of the evaporator Valves control the flow of liquids through these pipes into the glass tube. There is a line which supplies steam to the cavity between the inner and outer tubes and valves and a pressure gauge control the flow of this steam. As part of the feed vaporizes it exits the tube and enters a cyclone separator which separates the vapor and sends it to the condenser and sends the liquid to the product tank to be collected. The product line is also equipped with valves to control the flow and collection of liquid. The vapor that enters the condenser is condensed and sent to the condensate tank to be collected. The steam that exits the outer tube goes into a steam trap where it condenses and this condensate is constantly drained to avoid buildup.

Operating the climbing film evaporator

The main objective of this experiment was to determine the effect of varying the feed flow rate and steam pressure on the performance of the evaporator which is evaluated by heat transfer coefficients and the concentration of glycerol in the product solution.

The climbing film evaporator present in Goddard Hall 116 was used to conduct this experiment in two trials. During the first run the operating steam pressure was maintained constant at a certain value and the feed flow rates were varied. During the second trial the steam pressure was varied and same feed flow rates, as trial 1, were used again. Since the maximum steam pressure available in the Goddard Hall evaporator is 25 psig, operating pressure was always maintained below this value.

Procedure

1. Opened the valves connected to the steam supply line to drain any water that might have condensed and could possibly skew our data. Once the water had been drained the valve was closed.
2. Recorded the values of room temperature, steam temperature and atmospheric pressure.
3. Took a sample out of the feed tank and used the density meter provided to measure the specific gravity and hence the percentage glycerol of the feed solution. (This step was carried out just to verify t he composition of the feed solution which is indicated to be 10%).
4. Weighed the steam condensate collection bucket, the product solution bucket and the condenser solution bucket and placed them under their respective tanks.
5. Opened the valve on the cooling water line for the condenser to start the supply of cooling water to the process.
6. Opened the valve on the steam supply line to start pumping steam into the evaporator's outer glass tube. Selected a pressure of steam (5psig) and let it be constant for the rest of the experiment.
7. Turned on the pump and set a volumetric flow rate value (120 ml/min in the digital flow meter). Once the process had started we could observe steam going into the glass evaporator tube and the feed solution being pumped into the inner tube.
8. To ensure that the evaporator actually works smoothly and ensure that the process attains steady state a couple of runs were made without taking any data and analysis of the product and condenser solutions collected. We let the process run for around ten minutes to ensure that operating steam pressure, feed flow rate etc were constant throughout the runs.
9. After ensuring the process runs smoothly we started collecting data. Waited for a while to let the system attain steady state. (This time is variable for each flow rate and steam pressure but can be approximated to a minimum of 20 minutes for each run).
10. Started draining the product solution by opening the valve at the base of the product collection line. After completely draining the product reservoir, the valve was closed again for a period of two minutes and product was allowed to collect in the product tank. This interval was timed using a stop watch and the end of two minutes the valve was opened again to drain the collected product into an empty bucket that had already been weighed.
11. Weighed this product to determine the mass flow rate of the product.

12. Drained the condenser solution by opening the valve at the base of the condenser line. After draining the valve was closed again and an interval of two minutes was timed to collect condenser solution over. At the end of two minutes the valve was opened and water collected in the condenser reservoir was drained into a pre-weighed empty bucket.
13. Weighed this condenser solution to determine the condenser mass flow rate.
14. Collected the steam condensate over a period of two minutes and weighed it to determine the steam flow rate into the process.
15. Took a sample from the hot product solution into a beaker and immersed it into an ice bath to be cooled to ambient (room temperature).
16. Measured the specific gravity of this cooled sample using the density meter and determined the corresponding percentage of glycerol in the product from the water- glycerol solution specific gravity chart provided.
17. Repeated the above process for three other feed flow rates (200 ml/min, 300 ml/min and 400 ml/min) keeping everything else constant.

Day 2

1. Repeated the entire process above at a constant steam pressure of 10 psig and four different flow rates of (120, 200, 300 and 400 ml/min).

Shutdown Procedure

To ensure the safe shut down of the apparatus make sure the following steps are taken:

- Close the feed supply line by closing the valve it is equipped with.
- Close the steam control valve and all the other steam valves.
- Stop the feed pump.
- Open all valves at the bottom of product and condenser lines to drain any excess liquid.
- Leave the cooling water running even after the process has been shut down

Safety Precautions

In order to maintain a safe working environment in the laboratory the following safety precautions were taken:

- Check that all the valves are working properly and no air or water inlets/ outlets are blocked.
- Ensure that the pump and the flow meter are functioning correctly.
- Ensure that the steam trap is functioning correctly in order to avoid any steam or hot condensate being trapped.
- The maximum steam pressure that can be supplied to the evaporator is 25 psig. Ensure that the operating steam pressure never exceeds this value.
- Since this is a very energy intensive experiment a lot of heat is lost to the environment through the apparatus. Consequently a lot of the equipment gets extremely hot. Ensure that gloves are worn at all times when touching such equipment and to avoid contact with hot surfaces as much as possible.

- The steam condensate and the product solution both exit the apparatus at extremely high temperatures. Wear thick gloves whenever collecting these two liquids. Be very careful during the collection since the hot liquid or steam can splash and cause burns and injuries.
- Leave the cooling water flow on even after closing the steam line and feed supply line. Cooling water cools down the apparatus after the experiment is over to ensure that there is no overheating of equipment causing potential damage or safety concerns.
- Wear hard hats and goggles and appropriate lab safety equipment at all times in the lab.

Theory behind the calculations

Mass Balance Calculations

One of the major objectives of this experiment was to understand how the feed flow rate affects the final concentration of the product solution. In order to estimate the affect of varying feed flow rate on the other variables and parameters of the climbing film evaporator mass balance was calculated on the evaporator and the relevant theory and procedures are outlined below.

The structure of the climbing film evaporator can be basically broken down into a simple block diagram showing the major streams going in and out of the evaporator. The basic diagram for the evaporator looks as follows:

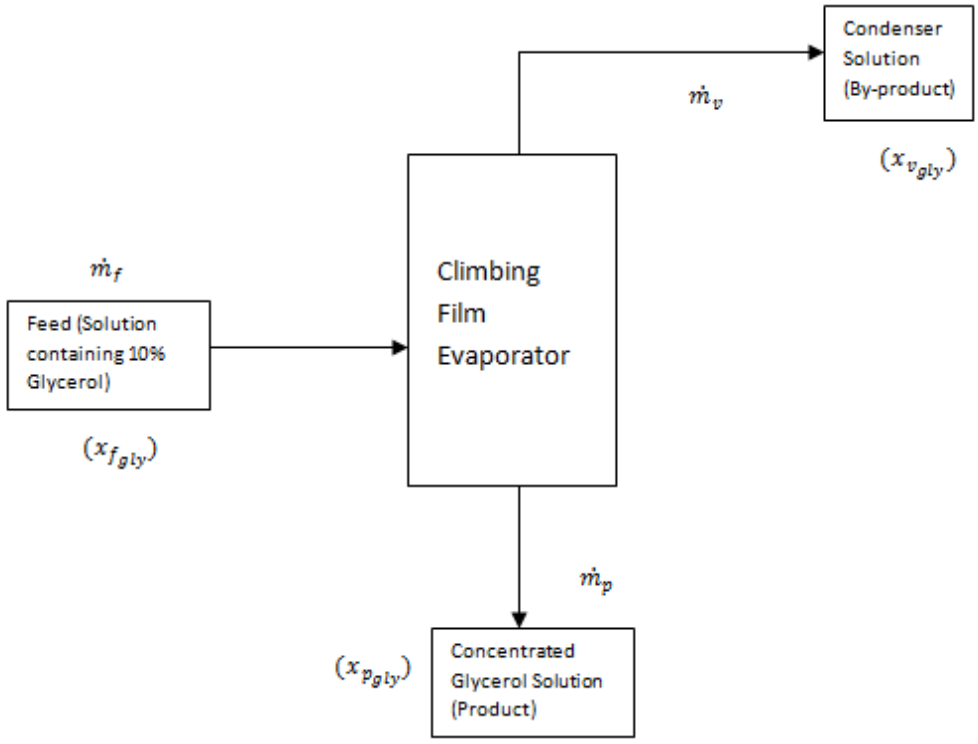


Figure 6: Block diagram of the evaporator (Mass Balance)

In order to calculate the mass balance we need the mass flow rates of all the inlet and outlet streams. Since we recorded the feed flow rate using a flow meter it is a volumetric flow rate with units (ml/min) and need to be converted to mass flow rate. Since the solution is a mixture of water and glycerol its

density would be a combination of the densities of water and glycerol. The equation below was used to perform this conversion.

$$(\rho_{feed}) = (\rho_{Gly} * x_{f_{gly}}) + (\rho_w * x_{f_w}) \quad (1)$$

Where:

$\rho_{feed} = Average\ density\ of\ the\ feed\ solution\ \left(\frac{g}{cm^3}\right)$

$\rho_{Gly} = Density\ of\ glycerol\ \left(\frac{g}{cm^3}\right)$

$x_{f_{gly}} = Weight\ fracion\ of\ glycerol\ in\ the\ solution$

$\rho_w = Density\ of\ water \left(\frac{g}{cm^3}\right)$

$x_{f_w} = Weight\ fracion\ of\ glycerol\ in\ the\ solution$

The next step was to convert the volumetric flow rate into mass flow rate as shown in Equation 2 below.

$$\dot{m}_f = \dot{V} * \rho_{feed} \quad (2)$$

Where:

$\dot{m}_f = Mass\ Flow\ Rate\ of\ the\ Feed\ \left(\frac{kg}{h}\right)$

$\dot{V} = Volumetric\ Flow\ Rate\ of\ the\ Feed\ (\frac{ml}{min})$

For the purpose of calculating mass balance we are going to assume that the condenser solution flow rate that we measured was accurate and use those values to calculate the mass balance. We assumed that the product flow rate we measured was not quite accurate owing to the fact that we collected it only over a period of 2 minutes and the fact that the sloshing over of the liquid might have been very erratic and not uniform (it might have sloshed over a lot of solution consecutively and the next minute there might have been very low amount of liquid. We have to take an average of these to find the correct flow rate and for that we should have collected the product over a longer period of time (maybe around 4-5 minutes).

Keeping this in mind we calculated the product flow rate with Equation [1] given below.

$$\dot{m}_P = \dot{m}_f - \dot{m}_v \quad (3)$$

Where:

$\dot{m}_P = Mass\ Flow\ Rate\ of\ the\ product\ \left(\frac{kg}{h}\right)$

$\dot{m}_v = Mass\ Flow\ Rate\ of\ the\ condenser\ solution \left(\frac{kg}{h}\right)$

Using similar balances as Equation 3 above us calculated the concentration of glycerol and water in the condenser and product solutions [1].

$$\dot{m}_{p_{gly}} = \dot{m}_{f_{gly}} - \dot{m}_{v_{gly}} \quad (4)$$

Where:

$\dot{m}_{p_{gly}} = Mass\ Flow\ Rate\ of\ the\ glycerol\ in\ the\ product\ \left(\frac{kg}{h}\right)$

$\dot{m}_{f_{gly}} = Mass\ Flow\ Rate\ of\ the\ glycerol\ in\ the\ feed\ \left(\frac{kg}{h}\right)$

$\dot{m}_{v_{gly}} = Mass\ Flow\ Rate\ of\ the\ glycerol\ in\ the\ vapor\ \left(\frac{kg}{h}\right)$

And similarly calculated the mass balance on water within the process [1]

$$\dot{m}_{p_W} = \dot{m}_{f_W} - \dot{m}_{v_W} \quad (5)$$

Where:

$\dot{m}_{p_W} = Mass\ Flow\ Rate\ of\ the\ water\ in\ the\ product\ stream\ \left(\frac{kg}{h}\right)$

$\dot{m}_{f_W} = Mass\ Flow\ Rate\ of\ the\ glycerol\ in\ the\ feed\ stream\ \left(\frac{kg}{h}\right)$

$\dot{m}_{v_W} = Mass\ Flow\ Rate\ of\ the\ glycerol\ in\ the\ vapor\ stream\ \left(\frac{kg}{h}\right)$

The next step was to calculate the theoretical percentage of glycerol (as predicted by theory) in the product solution to compare it to the experimental results obtained.

$$\dot{m}_f * x_{f_{gly}} = \dot{m}_p * x_{p_{gly}} \quad (6)$$

Energy Balance Calculations

The next step is to calculate the energy balance to see how much energy is utilized by the evaporator to vaporize water and concentrate glycerol and how much of the energy supplied is lost to the environment. This will not only help us to calculate the energy loss but also help us evaluate the economy and efficiency of the climbing film evaporator. This information will help us weigh the performance of the evaporator against the energy losses and help us decide whether the process is economically feasible or not.

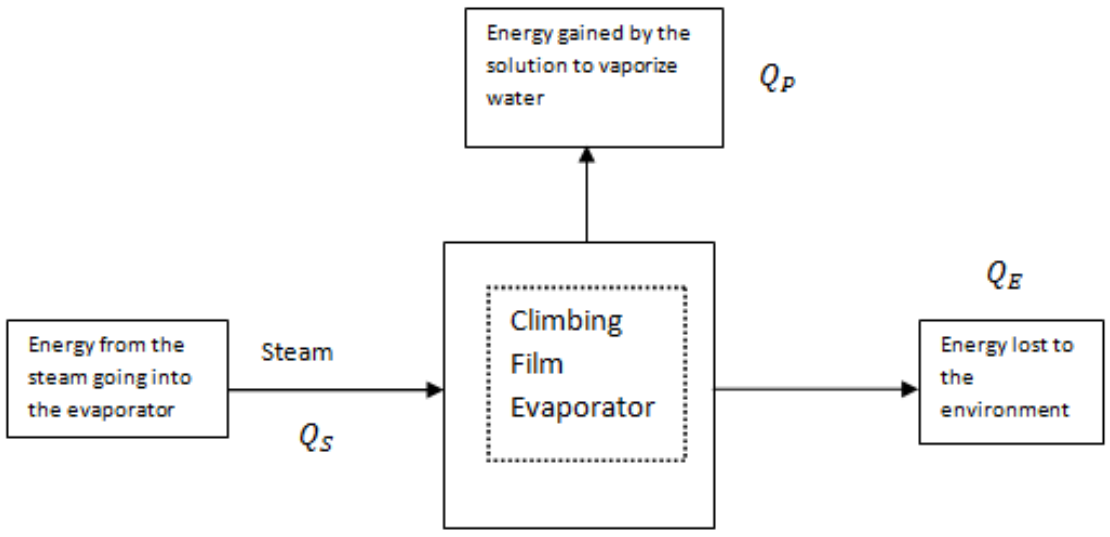


Figure 7: Block diagram of the CFE for energy balance calculations

The heat lost by the steam is given by the following relations [1]:

$$Q_S = \dot{m}_s \lambda_s \quad (7)$$

Where:

$\dot{m}_s = \text{Condensate mass flow rate } (\frac{kg}{sec})$

$\lambda_s = Latent\ Heat\ of\ vaporization\ of\ the\ steam\ \left(\frac{kJ}{kg}\right)$

$$Q_S = Q_p + Q_E \quad (8)$$

$$Q_S = \dot{m}_v \lambda + \dot{m}_f C_{p_f} \Delta T + Q_E \quad (9)$$

Where:

$Q_E = Energy\ losses\ to\ the\ environment$ (W)

$\lambda = Latent\ Heat\ Capacity\ of\ the\ solution\ \left(\frac{kJ}{kg}\right)$

$C_{p_f} = Average\ Specific\ Heat\ Capacity\ of\ the\ feed\ solution\ (\frac{J}{kg * K})$

$\Delta T = \left(T_{boiling} - T_{feed}\right)\ (K)$

In equation 8 given above the latent heat of vaporization (λ) is not the latent heat of condenser solution as we would expect it to be but is rather the latent heat for the feed solution. This latent heat is not constant and varies as the concentration of glycerol (any solute) in water varies. The values of this hence not taken to be 2769 kJ/kg but rather found out to be 1600 kJ/kg as given by Pacheco and Frioni [4].
(The heat gained by the process was then calculated using the Equation 10 [1] below:

$$Q_p = [\dot{m}_f C_{p_f} * \left(T_b - T_f\right)] + [\dot{m}_v \lambda_f] \quad (10)$$

Where:

$Q_p = Energy\ gained\ by\ the\ process\ (W)$

$\lambda_f = Latent\ Heat\ Capacity\ of\ the\ feed\ solution\ \left(\frac{kJ}{kg}\right)$

The heat lost to the environment was calculated using the overall energy balance for the evaporator given in Equation 11[1].

$$Q_S = Q_p + Q_E \quad (11)$$

$$Q_E = Q_S - Q_p$$

Heat Transfer Coefficient Calculations

Calculating the outer heat transfer coefficient

In the climbing film evaporator energy is constantly being exchanged between the steam and the process through convection and phase change between glass and the atmosphere and the glass and the feed solution. The overall heat transfer coefficient is the ability of a series of resistive materials or boundaries to transfer heat [1].

The overall heat transfer coefficient takes into account the individual heat transfer coefficients of each stream and the resistance of the pipe material. It can be calculated as the reciprocal of the sum of a series of thermal resistances such as:

$$Q_S = Q_p + Q_E \tag{11}$$

$$\frac{1}{U_A} = \frac{1}{h_i * A_i} + \frac{q}{k_q * A} + \frac{1}{h_o * A_o} \tag{12}$$

For the process of concentrating glycerol using the climbing film evaporator there is an outer overall heat transfer coefficient which is given by the expression [1]:

$$Q_E = U_{O_E} * A_O * \Delta T_{LM} \tag{13}$$

Where:
$U_{O_E} = Outer\ heat\ transfer\ coefficient\ (W/(m^2.K)$
$A_O\ = Outer\ surface\ area\ of\ the\ evaporator\ (m^2)$
$\Delta T_{LM} = is\ the\ Log\ mean\ temperature\ difference\ (\ which\ in\ this\ case\ is\ just\ (T_s - T_a))$

Using the already calculated values the energy lost to the environment from Equation 11 we can rearrange Equation 13 [1]:

$$U_{O_E} = \frac{Q_E}{A_O * (T_s - T_a)} \tag{14}$$

Calculating the inner heat transfer coefficient

In case of the evaporator process the inner heat transfer can be calculated in two ways:

- Consider the entire tube as a whole and neglect the phase change happening in the tube and consider the driving force to be between the feed inlet temperature and the outlet temperature
- Break down the tube into two processes
    - One where the feed gets heated and the height of the evaporator is taken to be the height at which the glycerol solution starts boiling. The heat transfer coefficient using these conditions is called the lower overall heat transfer coefficient.

- And the rest of the process where the water starts to evaporate (the height is considered to be the total height minus the boiling height) is used to calculate the upper heat transfer coefficient.

$$Q_P = U_l * A_l * \Delta T_{LM} + U_u * A_u * \Delta T_{LM} \quad (15)$$

Where:
$U_l = Lower\ heat\ transfer\ coefficient\ for\ the\ process\ (W/(m^2.K)$
$A_l = Area\ of\ the\ lower\ part\ of\ the\ tube\ (m^2)$
$\Delta T_{LM} = is\ the\ Log\ mean\ temperature\ difference\ (K)$
$U_u = Upper\ heat\ transfer\ coefficient\ for\ the\ process\ (W/(m^2.K)$
$A_u = Area\ of\ the\ upper\ part\ of\ the\ tube\ (m^2)$

We can break down the inner tube into lower and upper tubes and the heat gained by the process also gets divided in a similar manner. The heat gained by the process in the lower part of the tube is given by Equation 16 [1].

$$Q_l = \dot{m}_f * C_p * (T_b - T_f) \quad (16)$$

Where:
$Q_l = Heat\ gained\ by\ the\ feed\ in\ the\ lower\ part\ of\ the\ tube\ (W)$

Once the heat gained has been calculated we can now utilize this value to calculate the inner heat transfer coefficient in the lower part of the tube:

$$Q_l = U_l * A_l * \Delta T_{LM} \quad (17)$$

Where:
$U_l = Lower\ heat\ transfer\ coefficient\ for\ the\ process\ (W/(m^2.K)$
$A_l = Area\ of\ the\ lower\ part\ of\ the\ tube\ (m^2)$
$\Delta T_{LM}[1] = is\ the\ Log\ mean\ temperature\ difference\ (K)$ which in this case is given by the equation below:

$$\Delta T_{LM} = \frac{[(T_s - T_f) - (T_s - T_b)]}{\ln\left(\frac{(T_s - T_f)}{T_s - T_b}\right)} \quad (18)$$

Where:
$T_s = Temperature\ of\ the\ steam\ (K)$
$T_f = Temperature\ of\ the\ feed\ (K)$
$T_b = Temperature\ at\ which\ the\ feed\ solution\ starts\ to\ boil\ (K)$

The heat gained by the process in the lower part of the tube is given by Equation 19 [1].

$$Q_u = \dot{m}_v * \lambda \quad (19)$$

Where:
$Q_u = Heat\ gained\ by\ the\ feed\ in\ the\ upper\ part\ of\ the\ tube\ (W)$
$\lambda_s = Latent\ heat\ of\ vaporization\ of\ steam$
And once this energy has been calculated we can easily determine the upper heat transfer coefficient by the relation below [1]:

$$Q_u = U_u * A_u * \Delta T_{LM} \quad (20)$$

$U_u = Upper\ heat\ transfer\ coefficient\ \ for\ the\ process\ (W/(m^2.K)$
$A_u = Area\ of\ the\ upper\ part\ of\ the\ tube\ (m^2)$
$\Delta T_{LM} = is\ the\ Log\ mean\ temperature\ difference\ (K)$ which in this case is given by the equation below:

$$\Delta T_{LM} = \frac{[(T_s - T_b) - (T_s - T_o)]}{\ln\left(\frac{(T_s - T_b)}{T_s - T_o}\right)} \quad (21)$$

Where:
$T_o = Temperature\ at\ which\ the\ mixture\ exits\ the\ evaporator\ (K)$

Capacity and Economy Calculations

In order to evaluate the performance and cost benefit analysis of using the climbing film evaporator as the unit operation for concentrating a glycerol solution we need to calculate the evaporator economy and capacity.

Equation 22 was used to calculate the evaporator capacity.

$$Capacity = \dot{m}_v \quad (22)$$

And evaporator economy was calculated using Equation 23.

$$Economy = \frac{Capacity}{\dot{m}_{steam}} \quad (23)$$

Part 2: Modeling the process on COMSOL

As discussed earlier it is extremely helpful to model unit operations. A climbing film evaporator is a typical distributed parameter system, characterized by its inputs, outputs and system states being dependent not only on time but also spatial position, up the height of the evaporator tube. For a rigorous description, it should be modeled by a set of partial differential equations in space and time [5]. We used COMSOL to model the evaporator and in this section we are going to discuss the steps that we

followed to create the COMSOL model for the climbing film evaporator using the data obtained in the laboratory.

1. First start COMSOL **Multiphysics 3.5a** and click **Multiphysics**.
2. In the Model Navigator, **select Axial Symmetry (2D)** from the **space dimension tab**.
3. From the **Application Mode list**, select **Chemical Engineering Module>Mass Transport> Convection and Diffusion**.
4. In the **dependent variables** edit field, type the concentration variables: **Cg**, **Cw** and **Cwv** and click **Add**.
5. Select again from the **Applications Modes list, Chemical Engineering Module>Energy Transport> Convection and Conduction.**
6. In **the Dependent Variables** edit field, type the temperature variable: **T** and click **Add**.
7. For the third time select the **Applications Modes list, Chemical Engineering Module>Energy Transport> Convection and Conduction**.
8. In the **Dependent Variables** edit field, type the temperature variable: **T2** and click **Add**.
9. Select **Lagrange-Quadratic** from the **Elements list** for all three modes
10. It should look like Figure 8. Click **OK**.

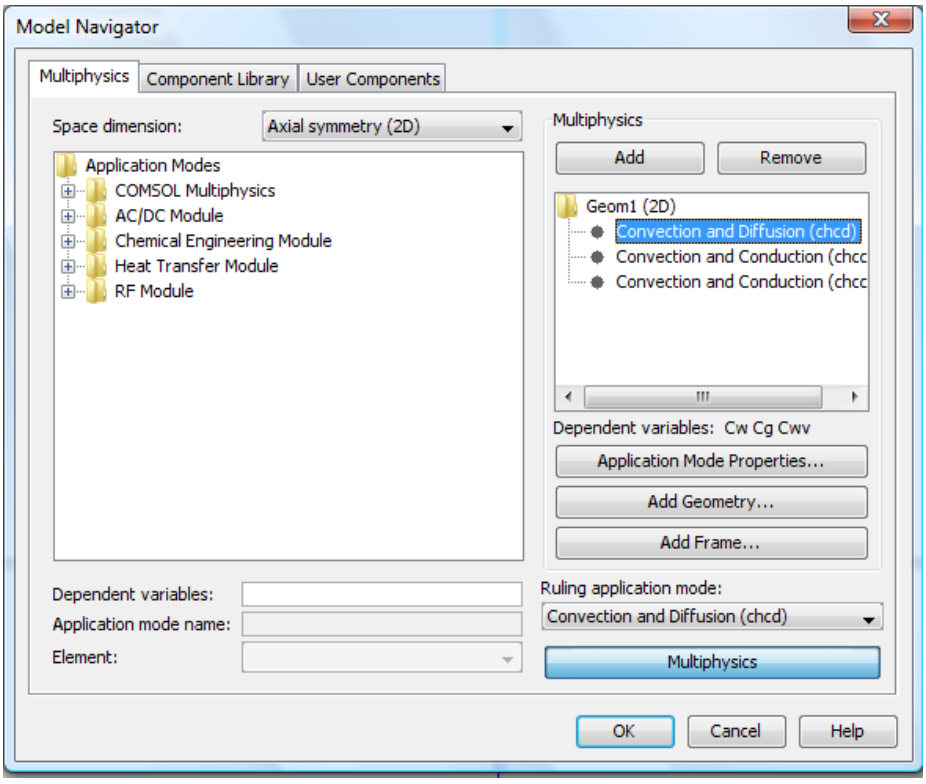


Figure 8: Model Navigator Screen

Where **Cg Cw** and **Cwv** correspond to the concentration of glycerol, water and water vapor inside the inner tube, **T** corresponds to the temperature inside the inner tube of the evaporator and **T2** is the temperature of the steam in the outer tube.

After clicking **Ok**, COMSOL will start and a blank screen with a dotted line called the axis of revolution will appear.

The next step is to draw the geometry of the climbing film evaporator.

Specifying Geometry

1. Click the **Draw Tab**>**Specify Object**>**Rectangle.** A box like Figure 9 should appear.

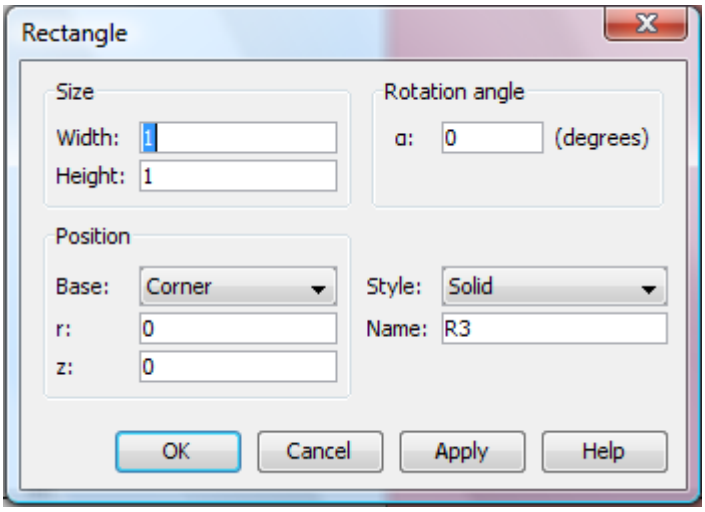


Figure 9: Draw Object Screen

2. In the **Width** edit field type the width of the inner tube as .0127.
3. In the **Height** edit field type the height of the evaporator as 2.74.
4. Click **Ok.**
5. Click again **Draw Tab**>**Specify Object**>**Rectangle.**
6. In the **Width** edit field type the width of the outer tube as 0.01905.
7. In the **Height** edit filed type the height of the outer tube as 2.74.
8. In the **r** edit field type the length of the inner radius as 0.0127
9. Click again **Draw Tab**>**Specify Object**>**Line.**
10. In the **r** edit field type the length of the line from 0 to 0.0127
11. In the **z** edit field type the height of the line from 0.9017 to 0.9017
12. Click again **Draw Tab**>**Specify Object**>**Line.**
13. In the **r** edit field type the length of the line from 0.0127 to 0.03175
14. In the **z** edit field type the height of the line from 0.9017 to 0.9017
15. Click **Ok.**

The geometry should look like Figure 10.

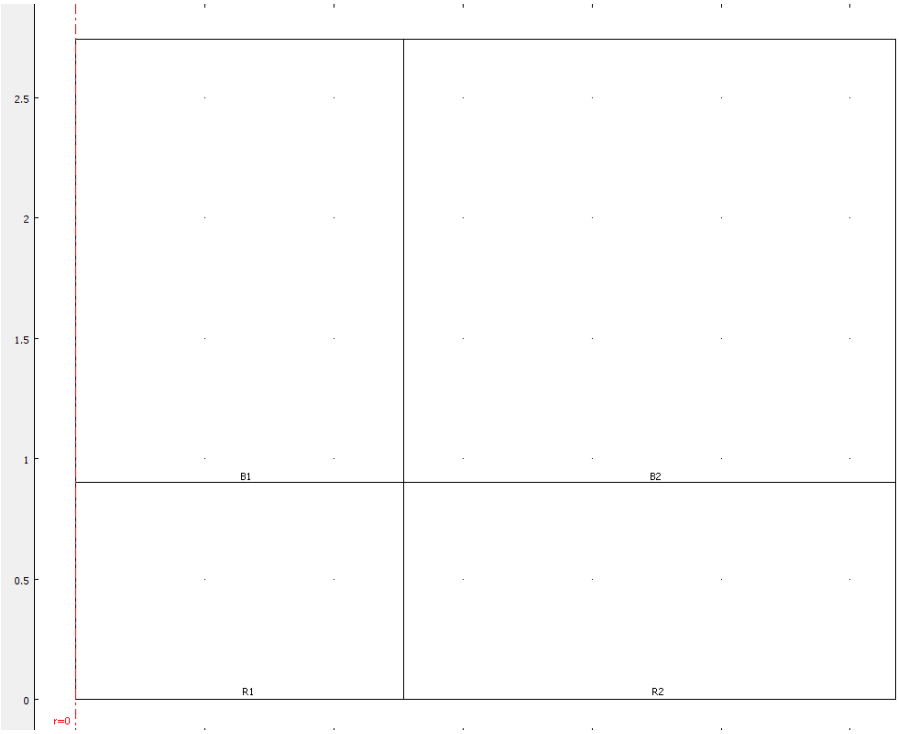


Figure 10: Geometry

The dimensions of the climbing film evaporator were based on the actual size of the climbing film evaporator located in the Unit Operations Laboratory in Goddard Hall. We measured the height of the apparatus to be 9 feet, the inner tube radius to be 1 inch and the outer tube radius to be 2.5 inches. All values were converted to meters to agree with COMSOL since it uses the metric system.

The line that divides the evaporator represents the height at which boiling starts occurring under this particular conditions and it was determined experimentally at 0.9017 m.

The next step is to setup the constants and global expression that were used to model the experiment.

Constant set up

1. - Click the **Options Tab**>**Constants** and input the constants seen in Figure 10.

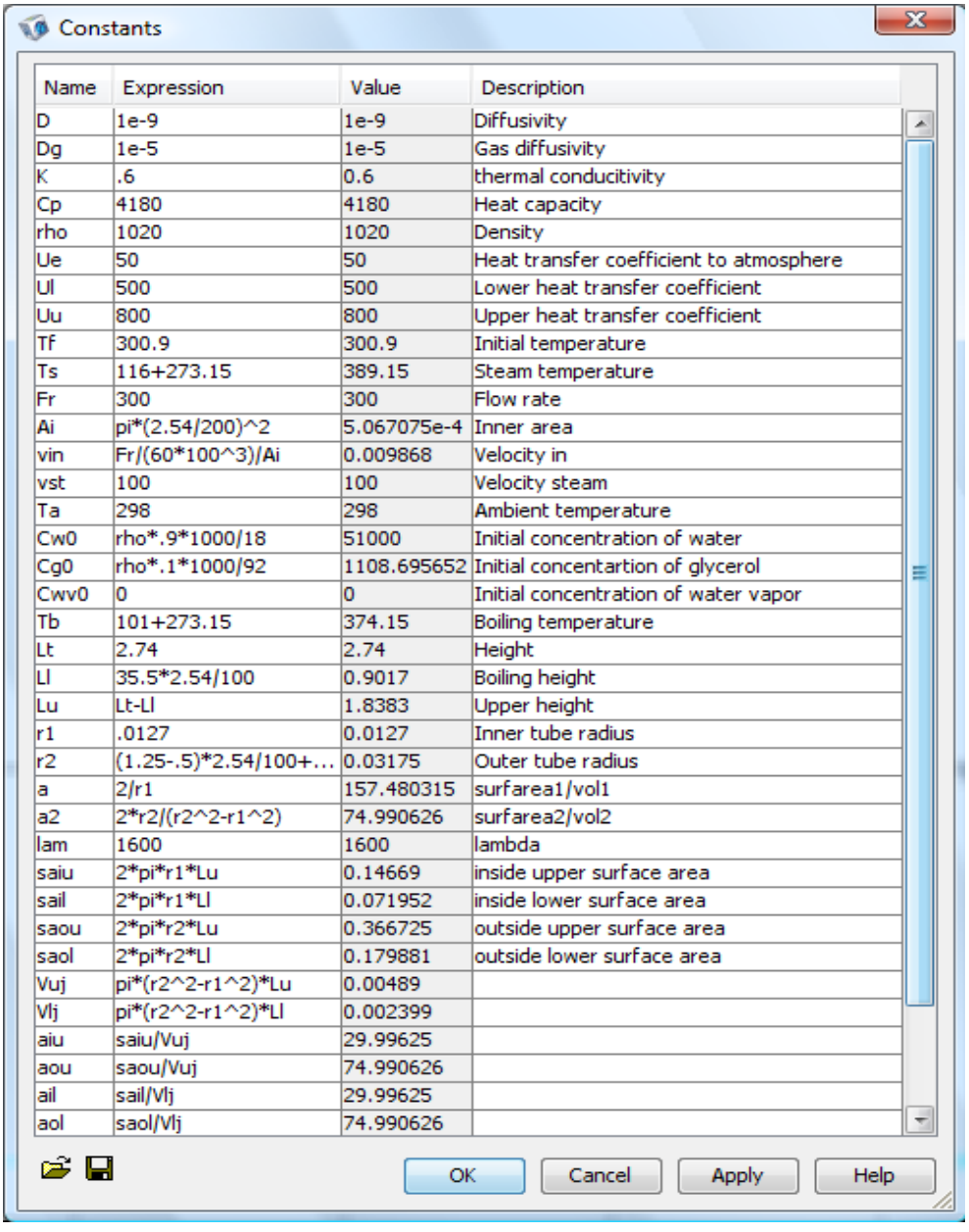


Figure 11: COMSOL Constants

Where D and Dg stand for water and gas diffusivity respectively and are in the units of m^2/s. K is the thermal conductivity constant in W/K*m. Cp is the heat capacity of water in J/m*K. Rho is the density of the mixture and was measured in kg/m^3. Ue is the heat transfer coefficient to the environment in W/ (m^2K). Ul and Uu are the lower and upper heat transfer coefficients for the evaporator in W/ (m^2K). Tf, Ts, Ta and Tb are the initial temperature of the mixture, the temperature of the steam in the outer side, the ambient temperature and the boiling temperature of the mixture all in degrees Kelvin. Fr is the volumetric flow rate in ml/min. vin is the initial velocity of the feed and was calculated using the volumetric flow rate and has units of m/s. vst is the initial velocity of the steam in m/s. Cw0, Cg0 and Cwv0 are the initial concentrations of water, glycerol and water vapor in mol/m^3 refer to appendix A for calculations. Lt is the total height of the evaporator in m. for purposes of modeling Ll is the height at which boiling occurs and Lu is the difference between the total height and the boiling height both in m. r1 and r2 are the inner tube radius and outer tube radius un m. lam is the heat of vaporization lambda in kJ/kg. saiu, sail, saou and saol are the inside upper surface area, inside lower surface area, outside upper surface area and outside lower surface area all in m^2. Vuj and Vlj are the upper and lower volumes in m^3.aiu, aou, ail and aol are surface areas over volume ratios used for modeling purposes all in m.

Expressions set up

1. Click the **Options Tab**>**Expressions>Global Expressions** and input the equations seen in Figure 12.

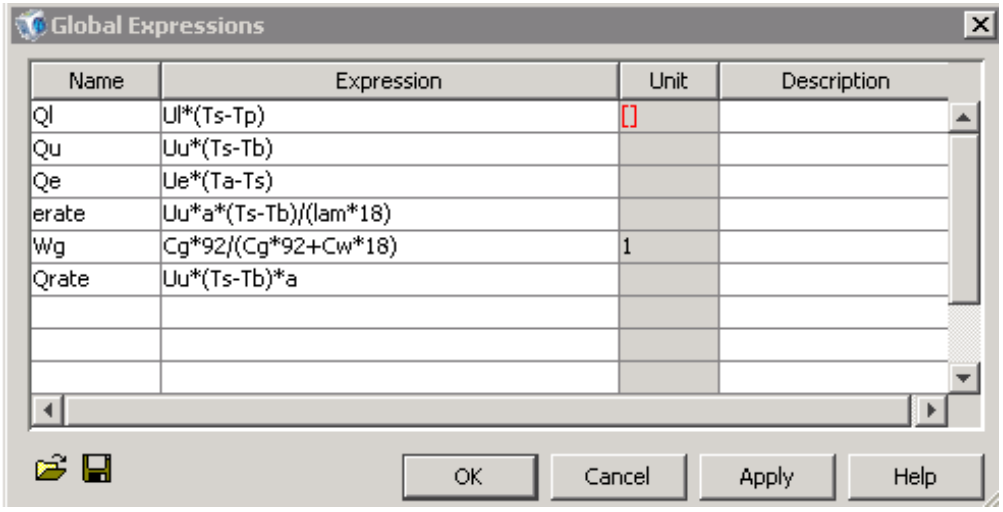


Figure 12: COMSOL Global Expressions

These expressions are used to calculate additional parameters that make the model function. Ql is the heat gained by the process from the steam in the lower part of the evaporator in watts. Qu is the heat gained by the process from the steam in the upper part of the evaporator in watts. Ql and Qu are necessary because we have to take into account that the upper and lower parts of the evaporator have different heat transfer coefficients due to the boiling that occurs inside the evaporator. Qe is the heat lost by the steam to the environment in watts. erate is the rate of vaporization of water inside the inner tube in mol/m^3*s. Wg is the weight fraction of glycerol inside the inner tube. Qrate is the rate of heat lost by the upper part of the inner tube to the steam.

The next step is to set the physics of the model which include the subdomain settings and the boundary conditions settings.

Subdomain settings

Convection and Diffusion (chcd)

1. Select from the toolbar **Multiphysics**> **1 Convection and Diffusion** (chcd).
2. Click from the toolbar **Physics**>**Subdomain Settings.** A box like Figure 13 should appear.

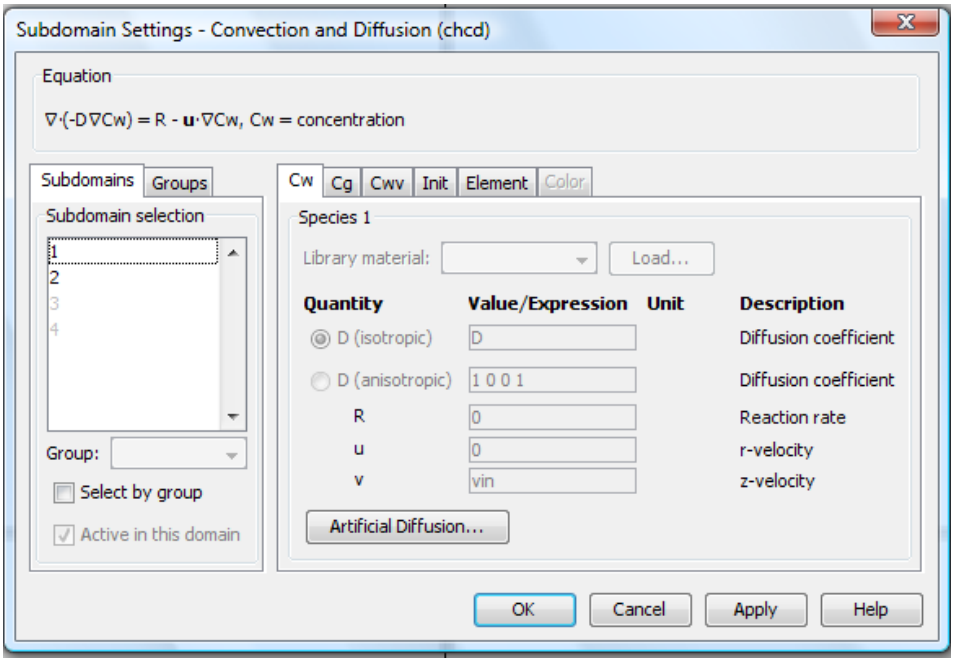


Figure 13: chcd Subdomain Settings

3. Select from the **subdomain selection list** subdomain **1.**
4. Select from the tabs **Cw.**
5. Type **D** in the diffusion coefficient box and **vin** in the z-velocity box.
6. Select from the tabs **Cg.**
7. Type **D** in the diffusion coefficient box and **vin** in the z-velocity box.
8. Select from the tabs **Cwv.**
9. Type **Dg** in the diffusion coefficient box and **vin** in the z-velocity box.
10. Select from the tabs **Init.**
11. Type **Cw0**, **Cg0** and **Cwv0** in the initial concentrations boxes.
12. Select from the **subdomain selection list** subdomain **2**.

13. Repeat steps 4 through 11. Additionally in step 5 type in the reaction rate box **–erate** and in step 9 type in the reaction rate box **erate**.

In order to simplify the model we are treating the evaporation of water as a reaction. We are assuming that water in the liquid phase is disappearing (evaporating) at a rate equal to Uu*a*(Ts-Tb)/(lam*18) and its appearing in the gas phase at the same rate. Hence the terms **erate** and **–erate** in the reaction rate boxes for Cw and Cwv.

Convection and Conduction (chcc)

1. Select from the toolbar **Multiphysics**> **2 Convection and Diffusion** (chcc).
2. Click from the toolbar **Physics**>**Subdomain Settings.** A box like Figure 14 should appear.

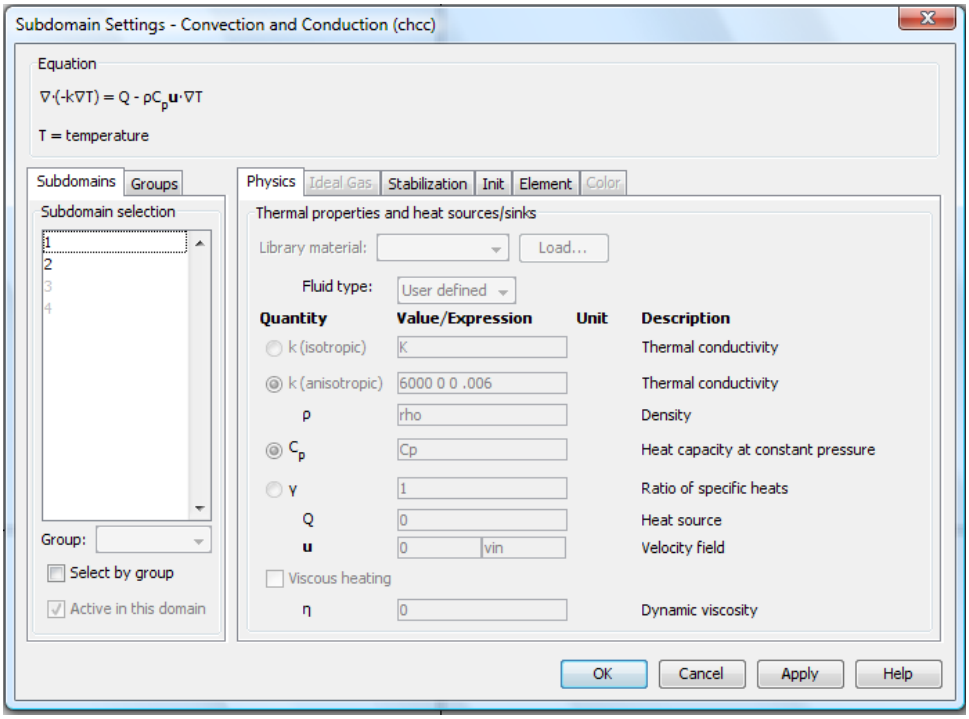


Figure 14: chcc Subdomain Settings

3. Select from the **subdomain selection list** subdomain **1.**
4. Select from the tabs **Physics.**
5. Click the **K (anisotropic) box** in the upper left corner of the square type **6000** and in the lower left right of the square type **.006** as the thermal conductivities. Type **rho** in the density box. Type **Cp** in the heat capacity box and type **vin** in the velocity field box.
6. Select from the tabs **init.**
7. Type **Tf** as the initial temperature.
8. Select from the **subdomain selection list** subdomain **2**.
9. Select from the tabs **Physics.**
10. Click the **K (anisotropic) box** in the upper left corner of the square type **6000** and in the lower left right of the square type **.006** as the thermal conductivities, type **rho in** the density box, type

Cp in the heat capacity box, type **–Qrate** in the heat source box and type **vin** in the velocity field box.

11. Select from the tabs **init.**
12. Type **Tb** as the initial temperature.
13. Click **OK**.

For modeling purposes we are assuming that the mixture will maintain the same temperature once it starts boiling, we achieve this by implementing the term –Qrate in the upper part of the evaporator. Additionally, we are only interested in simulating heat transfer inside the inner tube in the r direction hence K in the r direction is very big and K in the z direction is very small.

Convection and Conduction (chcc2)

14. Select from the toolbar **Multiphysics**> **3 Convection and Conduction** (chcc2).
15. Click from the toolbar **Physics**>**Subdomain Settings.** A box like Figure 14 should appear.

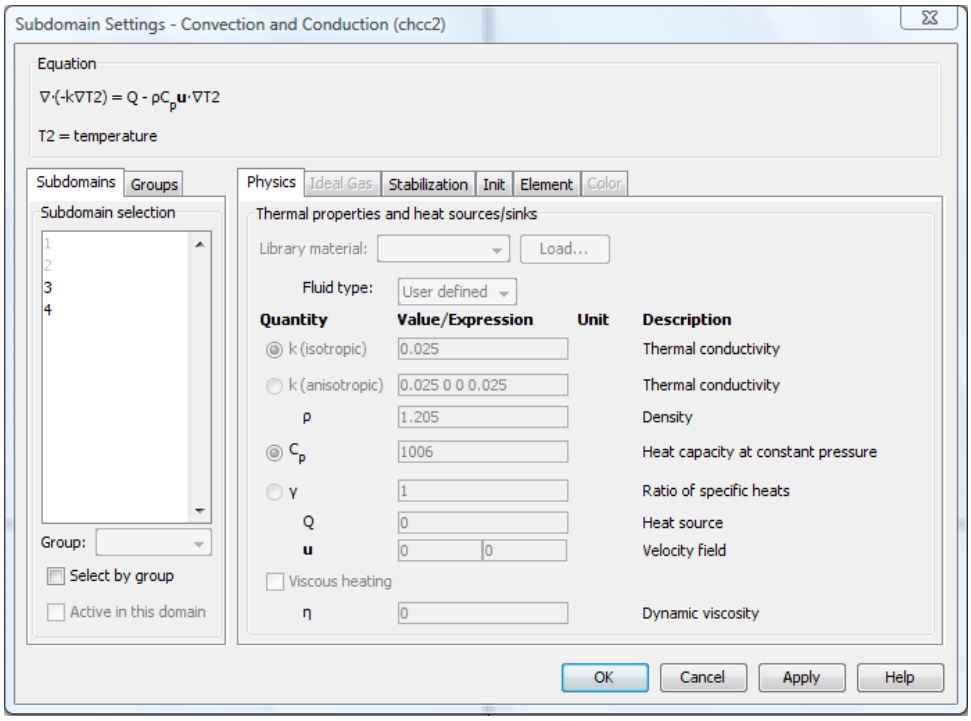


Figure 15: chcc2 Subdomain Settings

16. Select from the **subdomain selection list** subdomain **3.**
17. Select from the tabs **Physics.**
18. Click the **K (isotropic) box** input 25000 as the thermal conductivity, type **0.6** in the density box, type **2058** in the heat capacity box, type **Ql*ail-Qe*aol** in the heat source expression and type **vst** in the velocity field box.
19. Select from the tabs **init.**
20. Type **Ts** as the initial temperature.
21. Select from the **subdomain selection list** subdomain **4**.
22. Select from the tabs **Physics.**

23. Click the **K (isotropic) box** input 25000 as the thermal conductivity, type **0.6** in the density box, type **2058** in the heat capacity box, type **Qu*aiu-Qe*aou** in the heat source expression and type **vst** in the velocity field box.
24. Select from the tabs **init.**
25. Type **Ts** as the initial temperature.
26. Click **OK**.

For modeling purposes we are assuming that the steam running in the outer tube of the evaporator has uniform temperature. To model this behavior we decided to include the heat source terms in order to maintain the temperature of the steam as constant as possible.

Boundary settings

Convection and Diffusion (chcd)

1. Select from the toolbar **Multiphysics**> **1Convection and Diffusion** (chcd).
2. Click from the toolbar **Physics**>**Boundary Settings.** A box like Figure 16 should appear.

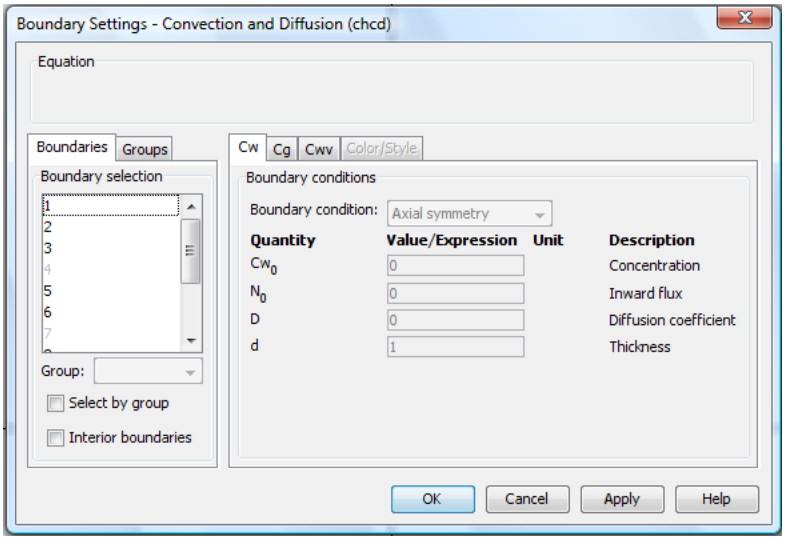


Figure 16: chcd Boundary Conditions

3. From the **boundary selection list** click boundary **1**.
4. From **the boundary condition drop down menu** select **Axial Symmetry**.
5. From the **boundary selection list** click boundary **2**.
6. From **the boundary condition drop down menu** select **Concentration**.
7. Click the **concentration** edit field and type **Cw0** as the initial concentration.
8. Select from the tabs **Cg**.
9. From **the boundary condition drop down menu** select **Concentration**.
10. Click the **concentration** edit field and type **Cg0** as the initial concentration.
11. Select from the tabs **Cwv**.
12. From **the boundary condition drop down menu** select **Concentration**.
13. Click the **concentration** edit field and type **Cwv0** as the initial concentration.
14. From the **boundary selection list** click boundary **3**.

15. From **the boundary condition drop down menu** select **Axial Symmetry**.
16. From the **boundary selection list** click boundary **5**.
17. From **the boundary condition drop down menu** select **Convective Flux**.
18. From the **boundary selection list** click boundary **6**.
19. From **the boundary condition drop down menu** select **Insulation/symmetry**.
20. From the **boundary selection list** click boundary **8**.
21. From **the boundary condition drop down menu** select **Insulation/symmetry**.
22. Click **OK**.

The boundary conditions represent the physical phenomena occurring in every side of the rectangle. For this particular case, the boundary conditions specified correspond to the mixture inside the evaporator. Boundaries 1 and 3 are located on the axis of symmetry and are specified as such. Boundary 2 is where the mixture enters the evaporator and is denoted as Concentration. Boundary 5 is the exit of the evaporator and is specified as Convective Flux. Boundaries 6 and 8 are the side of the evaporator that is in contact with the steam and are specified as insulation/symmetry.

Convection and Conduction (chcc)

1. Select from the toolbar **Multiphysics**> **2 Convection and Conduction** (chcc).
2. Click from the toolbar **Physics**>**Boundary Settings.** A box like Figure 17 should appear.

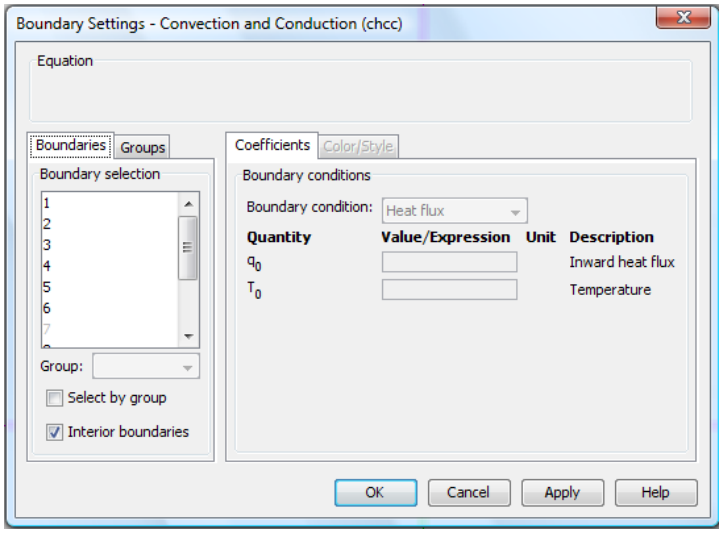


Figure 17: chcc Boundary Conditions

3. From the **boundary selection list** click boundary **1**.
4. From **the boundary condition drop down menu** select **Axial Symmetry**.
5. From the **boundary selection list** click boundary **2**.
6. From **the boundary condition drop down menu** select **Temperature**.
7. Select the **Temperature** edit field and type **Tf**.
8. From the **boundary selection list** click boundary **3**.
9. From **the boundary condition drop down menu** select **Axial Symmetry**.
10. From the **boundary selection list** click boundary **4**.

11. From **the boundary condition drop down menu** select **Continuity**.
12. From the **boundary selection list** click boundary **5**.
13. From **the boundary condition drop down menu** select **Convective Flux**.
14. From the **boundary selection list** click boundary **6**.
15. From **the boundary condition drop down menu** select **Heat Flux**.
16. Select the **Inward Heat Flux** edit field type in **Ql**.
17. From the **boundary selection list** click boundary **8**.
18. From **the boundary condition drop down menu** select **Heat Flux**.
19. Select the **Inward Heat Flux** edit field type in **Qu**.
20. Click **Ok**.

These boundary conditions correspond to the physical phenomena interacting with the temperature in the inner tube of the evaporator. Boundaries 1 and 3 are located on the axis of symmetry. Boundary 2 is the initial temperature of the mixture entering the evaporator. Boundary 4 represents the height at which boiling starts occurring. Boundary 5 is the temperature of the mixture exiting the evaporator. Boundaries 6 and 8 represent the interaction between the temperature of the mixture in the inner tube with the temperature of the steam running in the outer tube of the evaporator.

Convection and Conduction (chcc2)

1. Select from the toolbar **Multiphysics**> **3 Convection and Conduction** (chcc2).
2. Click from the toolbar **Physics**>**Boundary Settings.** A box like Figure 18 should appear.

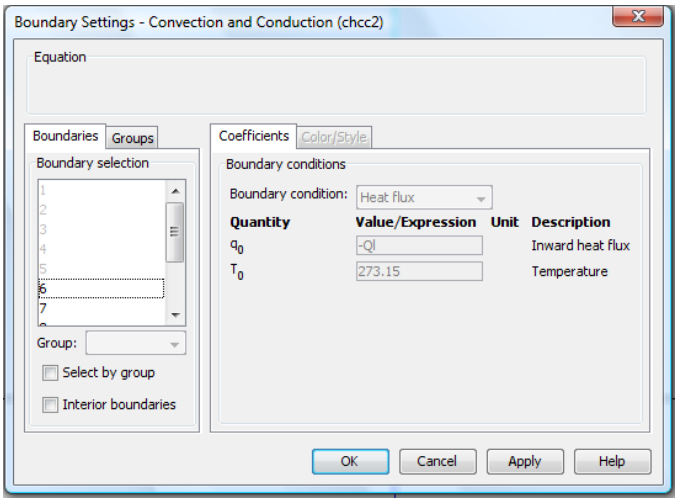


Figure 18: chcc2 Boundary Conditions

3. From the **boundary selection list** click boundary **6**.
4. From **the boundary condition drop down menu** select **Heat Flux**.
5. Select the **Inward Heat Flux** edit field type in -**Ql**.
6. From the **boundary selection list** click boundary **7**.
7. From **the boundary condition drop down menu** select **Temperature**.
8. Select the **Temperature** edit field and type **Ts**.

9. From the **boundary selection list** click boundary **8**.
10. From **the boundary condition drop down menu** select **Heat Flux**.
11. Select the **Inward Heat Flux** edit field type in -**Qu**.
12. From the **boundary selection list** click boundary **10**.
13. From **the boundary condition drop down menu** select **Convective Flux**.
14. From the **boundary selection list** click boundary **11**.
15. From **the boundary condition drop down menu** select **Heat Flux**.
16. Select the **Inward Heat Flux** edit field type in **Qe**.
17. From the **boundary selection list** click boundary **12**.
18. From **the boundary condition drop down menu** select **Heat Flux**.
19. Select the **Inward Heat Flux** edit field type in **Qe**.
20. Click **Ok**.

These boundary conditions correspond to the physical phenomena interacting with the temperature of the steam in the outer tube of the evaporator. Boundaries 6 and 8 represent the interaction between the temperature of the steam in the inner tube and the temperature of the mixture in the inner tube. Boundary 7 is the initial temperature if the steam. Boundary 10 is the temperature of the steam at the top of the tube. Boundaries 11 and 12 represent the heat lost of the steam to the environment.

Extrusion coupling values

In order to make the model work we have to define some extrusion coupling to help COMSOL calculate the initial temperature of the steam.

1. Click from the toolbar **Options**>**Extrusion Coupling Variables**>**Subdomain Extrusion Variables** a screen like Figure 19 should be prompted.

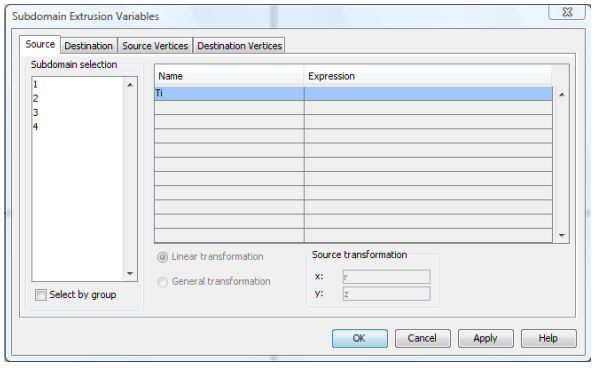


Figure 19: Extrusion Coupling Values

2. From the **Subdomain selection list** select subdomain **1.**
3. Under the **Name** edit field write the variable **Ti** and under the **Expression** edit field write the variable **T**.
4. Click the **Destination tab**.
5. From the **Subdomain selection list** check **3**.
6. Click the **Source Vertices** tab.
7. From **the Vertex selection list** select **1** and **2**.

8. Click the **Destination Vertices** tab.
9. From **the Vertex selection list** select **4** and **5**.
10. Click **Ok**.

Mesh Generation

With all the components of the model defined and specified, the only thing left to solve the model is to specify the mesh criteria.

1. From the toolbar select **Mesh**>**Refine Mesh** three times.
2. From the toolbar select **Solve**>**Solve Problem.**

Postprocessing

1. From the toolbar select **Postprocessing**> **Plot Parameters**. Click on the **Surface** tab and type **Wg** en the **Expression** edit field.

For other flow rates and pressures, all the steps listed above starting from specifying the geometry until postprocessing should be followed with differen input values for flow rate, steam pressure etc (as changes in these inputs will impact the outcome of the COMSOL model) to model and compare data with all the experimental runs that were made.

Results and Discussion

Part 1: Results for the Mass Balance on the Climbing Film Evaporator

The summarized results of both the runs (evaporator running at 5psig steam pressure and evaporator running at 10 psig steam pressure) to concentrate a glycerol solution using a climbing film evaporator are presented in Tables 1 and 2. Detailed sample calculations are shown in Appendix A.

Mass Flow Rate of steam(kg/h)	Feed Flow Rate of the condenser solution (kg/h)	Feed Flow of the product solution(kg/h)	% glycerol in the product solution(measured)
7.5	3.3	4.07	16
8.1	2.7	9.585	13
9	2.4	16.027	11
9.3	1.8	22.77	10.5

Table 1: Calculated values for input and output flow rates and percentage glycerol in product @ 5 psig

Mass Flow Rate of steam(kg/h)	Feed Flow Rate of the condenser solution (kg/h)	Feed Flow of the product solution(kg/h)	% glycerol in the product solution(measured)
7.8	4.2	3.17	17
9	3.9	8.385	14
9.6	3.3	15.127	12
9.9	2.7	21.870	11

Table 2: Calculated values for input and output flow rates and percentage glycerol in product @ 10 psig

The tabular results presented above help us compare certain numerical values and ascertain trends in the data right away. We can see that as the steam pressure increases the mass flow rate of the steam condensate increases, the reason for this being that at high pressures the latent heat of vaporization of steam is lower. Thus if the same amount of steam is supplied to the system at a higher pressure, less energy is required to condense it which results in a higher condensation rate and mass as indicated by the mass flow values of steam in Table 2. The tables also indicate that as the steam pressure increases the condenser solution flow rates and percentage glycerol in the product also increase.

An important thing to note here is that as the steam pressure increases the product solution flow rate decreases. Since the same amount of heat (and therefore same energy) is being supplied to the system at a higher pressure less energy is being used up, to condense steam and therefore more energy is available to be gained by the feed. As a result of this more water evaporates and from the feed increasing the flow rate of the condenser solution and decreasing the flow rate of the product solution. However, the above process dynamics also results in a higher concentration of glycerol in the product solution since a higher quantity of water vapor is generated and sent to the condenser leaving lesser water in the product stream. A graphical representation of the effect of higher steam pressure on the percentage glycerol in the product is shown below in Figure 3.

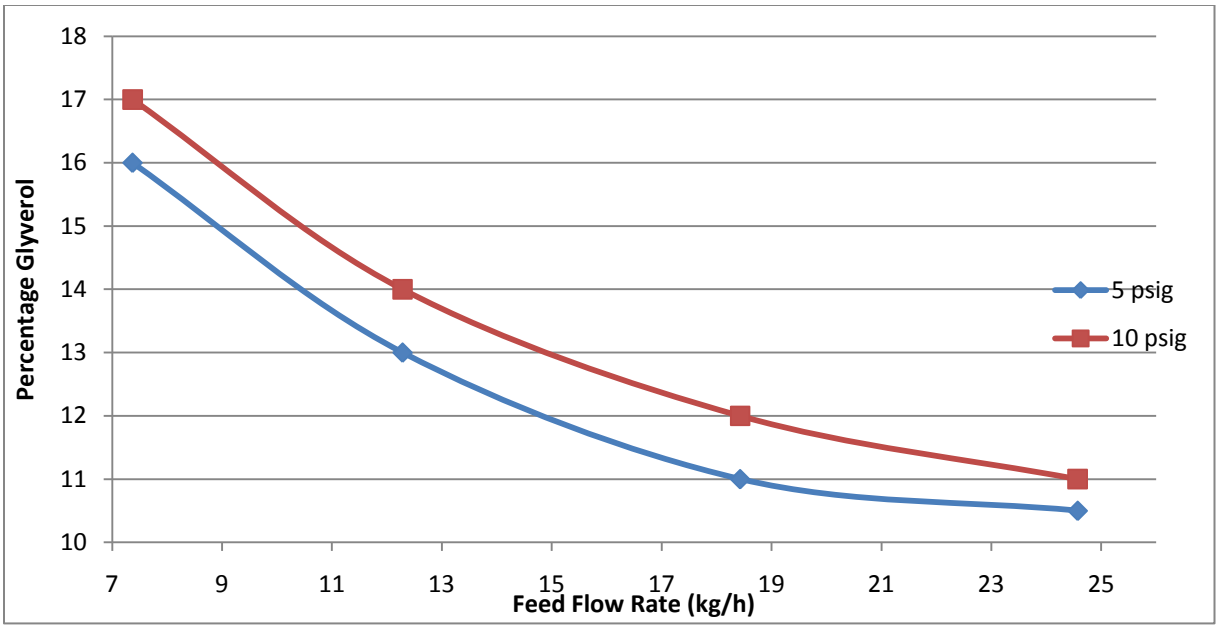


Figure 20: Experimental results comparing glycerol concentration in the product at different steam pressures

The graph above helps us compare the concentrations of glycerol in the product at different steam pressures and makes it very obvious that as the steam pressure is increased the concentration of glycerol in the product solution increases.

The main objective of this lab was to use the climbing film evaporator to increase the percentage of glycerol from 10 % in the feed solution to 16% or higher in the product solution.

Figure 20 makes it clear that at any given steam pressure when the feed flow rate is increased the percentage glycerol in the product decreases. This happens because when the steam pressure is held constant and feed flow rate is increased the contact time between the feed solution and the steam decreases resulting in less energy being transferred from the steam to the feed. As less heat energy is gained by the heat a lower amount of water (than should have actually evaporated had the flow rate not increased by a big margin) being evaporated and ending up in the product stream which gets diluted.

Heat given off by the steam is calculated by:

$$Q_S = \dot{m}_s \lambda_s$$

According to the results above for our given tube height and surface area we need to run the evaporator at a high steam pressure and low flow rates to obtain a more concentrated product.

Comparison to Theoretical Data

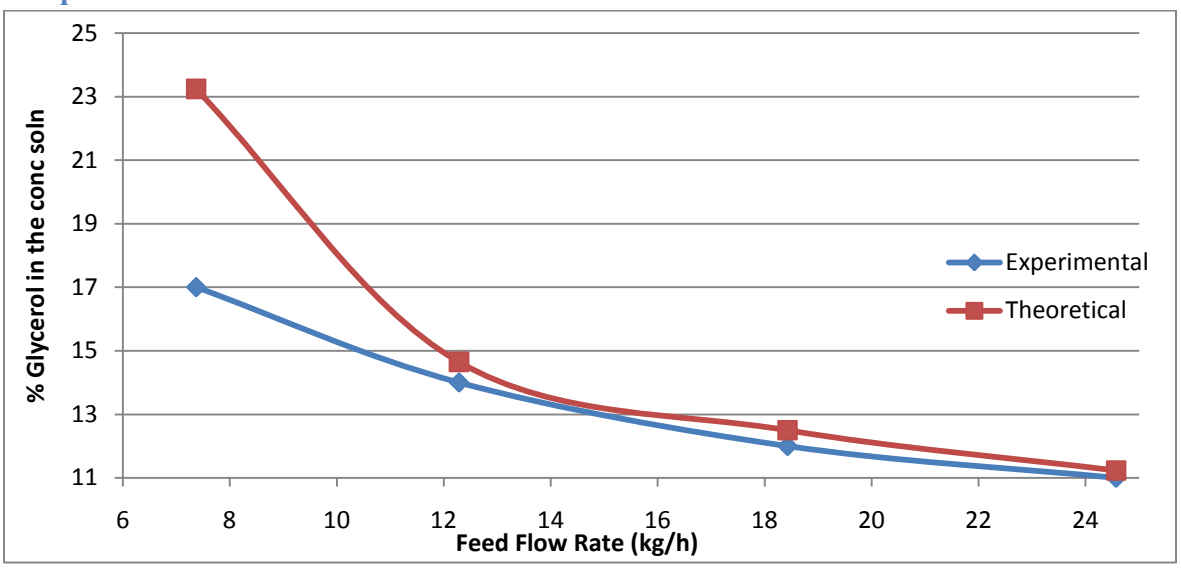


Figure 21: Comparison of experimental and theoretical percentage of glycerol in the product @10 psig

The experimental data obtained for the percentage of glycerol in the produt solution was compared to the theoretical values of percentage glycerol in the product. Detailed information and calculations regarding the theoretical values are given in Apppendix A. The graph above compares the theoretical and experimental values at a steam pressure of 10 psig. The trend for both experimental and theoretical curves seems to be the same i.e the percentage glycerol of the product goes down as the feed flow rate increases as stated earlier. However, the theoretical curve is slightyl higher than the experimental curve in most cases.This difference in values can be attributed to various experimental errors such as mistakes in reading the percentages off the specific gravity chart, mistakes in measuring the condensate flow rate etc. The first experimental data point at 17% is however much lower than the theoretical value of 23%. This large difference is most possibly due to the fact that the measurements at this flow rate were made before the evaporator systeam had attained steady state. If this is the case then the heating and vaporization in the tube had not yet become uniform resulting in uneven sloshing over of liquid into the product stream corrupting the data measurements. The solution for this problem is to wait longer for the system to come to steady state and/or figure out a method to estimate the timerequired for the system to reach steady state so that measurements are made at proper intervals.

Part 2: Results for the Energy Balance on the Climbing Film Evaporator

Steam Pressure (psig)	Feed Flow Rate (kg/h)	Heat gained by the process Qp(kW)	Heat loss to the environment QE (kW)
5	7.3710	2.059389	2.594778
5	12.285	2.18404	2.84246
5	18.427	2.548472	2.974472
5	24.570	2.775741	2.995426

10	7.3710	2.479497	2.321837
10	12.285	2.754716	2.785284
10	18.427	2.998741	2.910592
10	24.570	3.242766	2.851234

Table 3: Calculated values for heat gained by the process and heat lost to the environment at variable steam pressures.

The values in the table above help us to directly compare Q_p and Q_E at different steam pressures feed flow rates. We can see that as the feed flow rate increases the heat gained by the process Q_p increases indicating that as more solution is fed into the system more energy is gained by it as indicated by Figure 22 below.

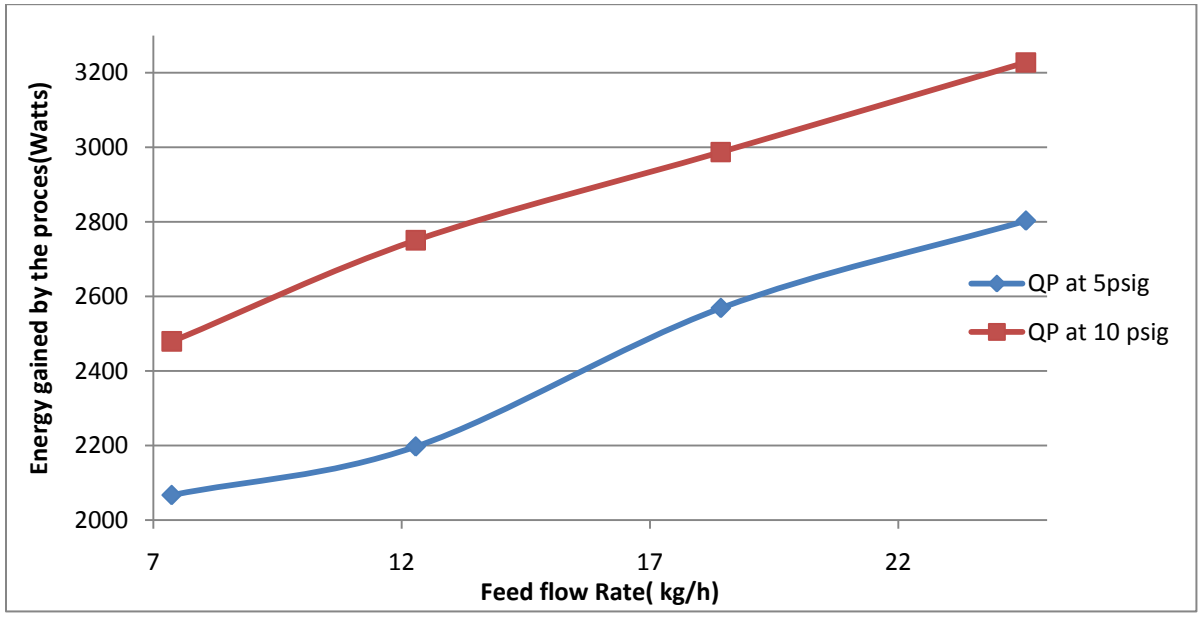


Figure 22: Trend followed by QP at variable steam pressures

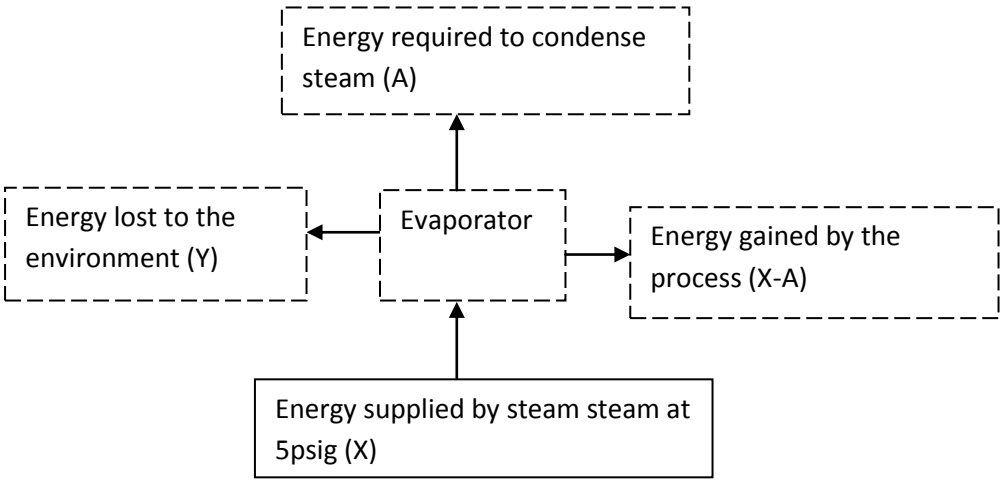
We can see an almost linear trend followed by the heat gained by the process as flow rate of the feed is increased. This trend is the same for both pressures which leads us to conclude that at any given pressure an increase in feed flow rate results in an increase in heat gained by the process. However, numerically this is not a huge increase. As it was stated earlier that when the feed flow rate increases vaporization of water does increase but not a lot since the residence time of the solution in the evaporator tube decreases. Similarly heat gained by the process does increase but not by a big margin since the residence time gets shorter.

Figure 22 also indicates that the heat gained by the process is higher at a higher steam pressure when similar feed flow rates are used at both pressures.

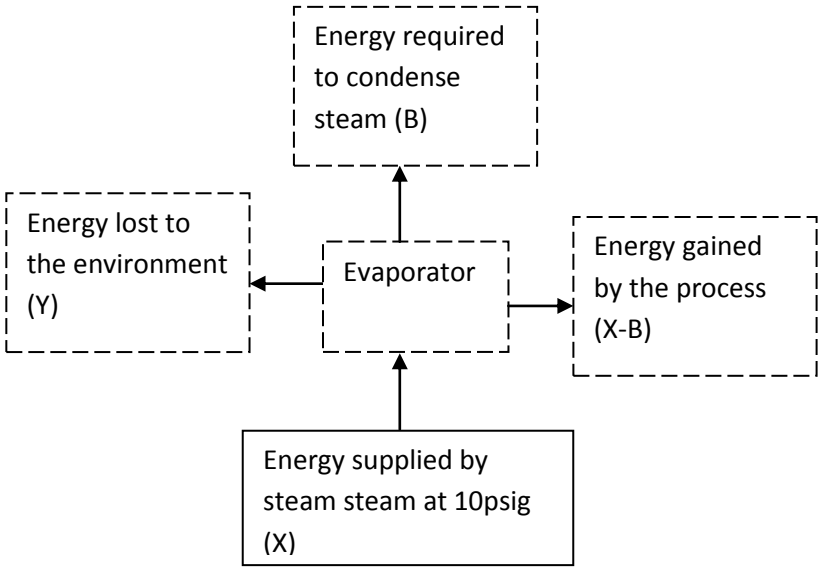
The explanation for this result is that as the same amount of steam is supplied to the system at a higher pressure, the latent heat of steam is lower. This results in more steam being converted into water using lesser amount of energy. This “extra” energy which is now not being utilized to condense steam is

instead gained by the process resulting in higher condensation rates for water vapor and more glycerol in the product solution.

This explanation is shown in a simplified energy balance diagram below.



Block Diagram 1: Block Diagram of Heat distribution at lower stream pressure



Block Diagram 2: Block Diagram of Heat distribution at higher stream pressure

Since B <A, the energy gained by the process (X-B) > (X-A). The above schematic gives a very simple breakdown of why the energy gained by the process is higher at higher steam pressures.

On the other hand as the feed flow rate is increased, the amount of steam condensing into water also increases indicating that a slightly larger amount of heat energy is being given off by the steam. As discussed earlier this increase in available energy results in an increase in the heat gained by the

process. Consequently the heat lost to the environment remains more or less constant since all the excess energy is channeled into the process.

Again as discussed above as the operating pressure of the steam being supplied is increased the energy gained by the process increases since steam at a higher pressure has a higher temperature and therefore more heat energy available. Therefore heat lost to the environment remains constant as indicated in Figure 23 below.

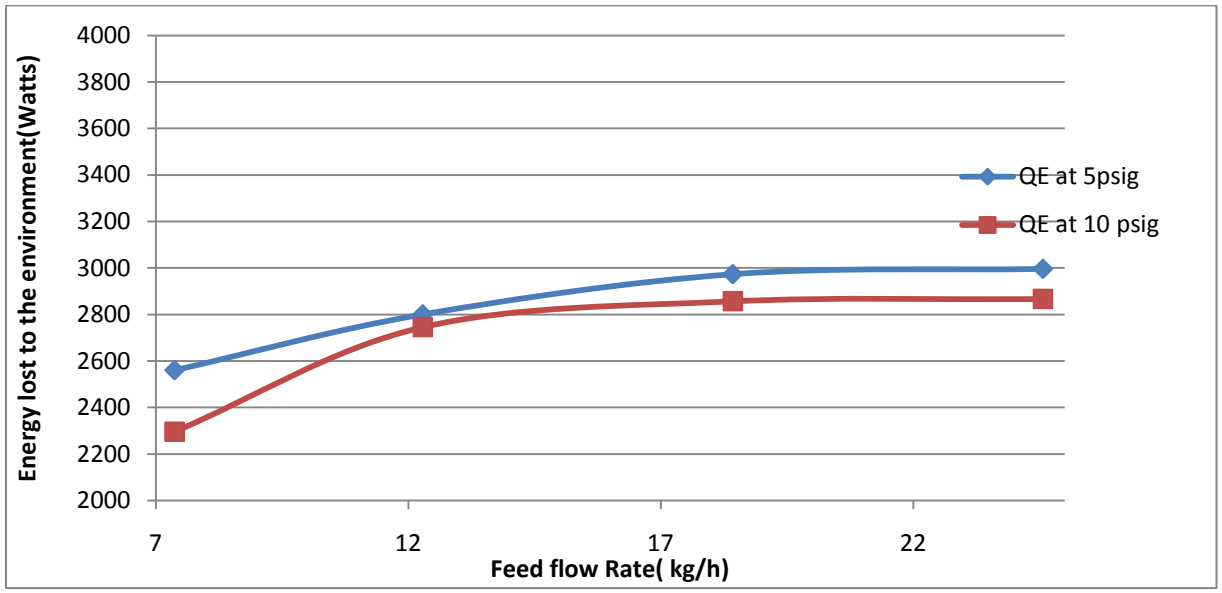


Figure 23: Trend followed by QE for variable steam pressure and feed flow rate

We can see in the graph above that QE at 10 psig and 5 psig are more or less straight lines and the only points which show an anomaly are at the flow rates of 120 ml/min which was the first run conducted for both the steam pressures. This anomaly can be explained by the possible error of the system not having attained steady state. When the system is not at steady state there is uneven heating and vaporization of water which results in erratic energy lost and energy utilized values. Possible human errors in the process of weighing the liquids collected could have also affected the energy balance calculations.

Another point to be noted about the graph above is that energy lost to the environment at 5 psig is slightly higher than QE at 10psig. Although this difference is negligible this should not have been the case in an ideal situation. The possible explanation for this error is the difference in ambient temperature (room temperature) while conducting the experiment. The run at 5 psig and 10 psig were conducted on different days which were at least 4 days apart. The energy loss being higher during the 5 psig run might mean that the ambient temperature on that day might have been slightly lower thereby affecting the energy balance between the evaporator and the air. Since we did not measure and take into account the ambient temperature in our calculations, we suggest that future users take room temperature measurements and verify if the ambient temperature does affect the results and to what extent.

Comparison of QP and QE

For a constant flow rate and constant steam pressure of (5psig) Q_E is higher than Q_P and is illustrated in the Figure 24 below.

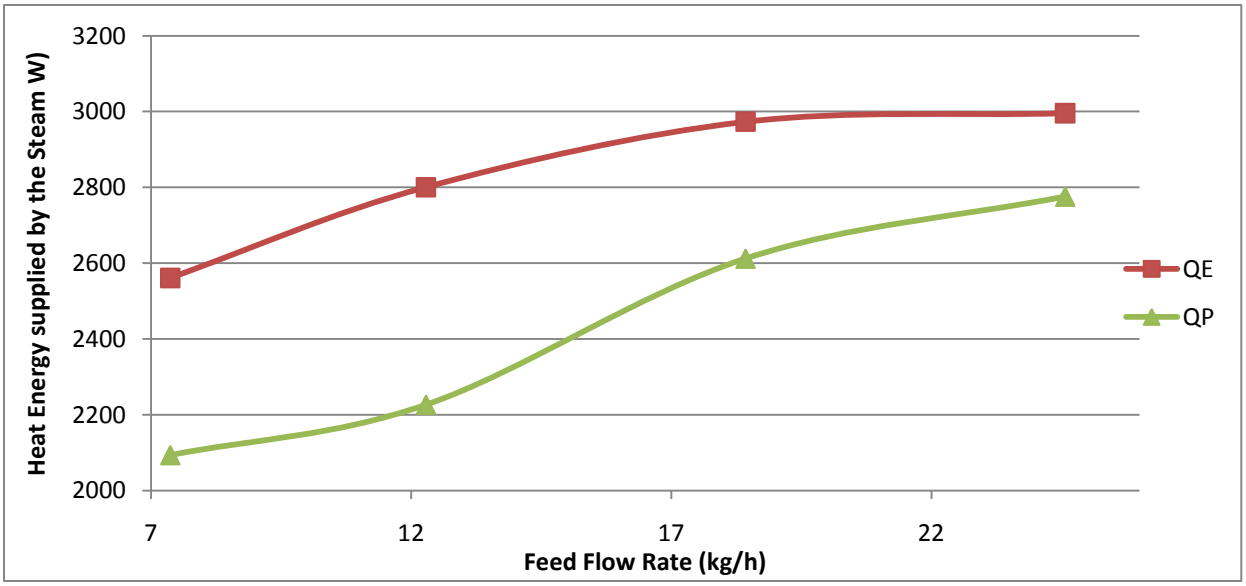


Figure 24: Comparison of QP and QE with varying flow rates at 5 psig

As we can see from the graph above that barring the first data point, the line representing QE is more or less straight and QP is increasing in a somewhat linear fashion. This result agrees with our previous deduction that QE remains constant and QP increases with an increase in feed flow rate.

However, at a steam pressure of 10 psig QP is higher than QE illustrated in Figure 25 below. The possible reason for this flip in QE and QP lines is because as the steam pressure increases, the heat energy available for the process increases whereas QE remains more or less stagnant irrespective of the feed flow rate or stream pressure. QE is the heat lost to the room or atmosphere by the hot steam and since we are not changing the flow rate of the steam or the insulation of the system, changes in other process variables should not affect the value of QE which is the case here. However, this excess heat energy goes into heating the feed and thus again the QP line clearly shows a more or less linear increase and also is higher than QE in this case as indicated in the Figure below.

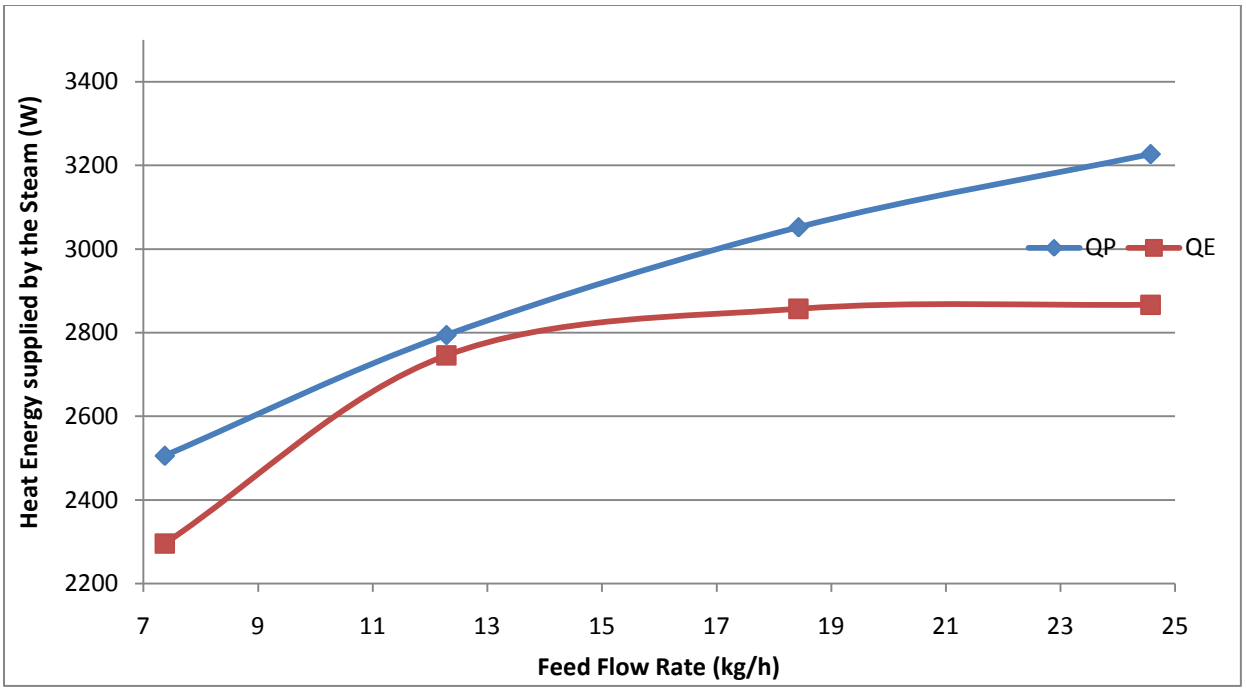


Figure 25: Comparison of QP and QE at 10 psig

Part 3: Overall Heat Transfer Coefficient results

One objective of the experiment was to determine how the glass and the atmosphere offer resistance to heat transfer from steam to the feed solution.

An outer heat transfer coefficient would be the measure of the resistance that the air and glass present to transfer of heat from the steam to the environment. This was calculated considering the entire tube as a whole and the results are presented in Figure 25 below.

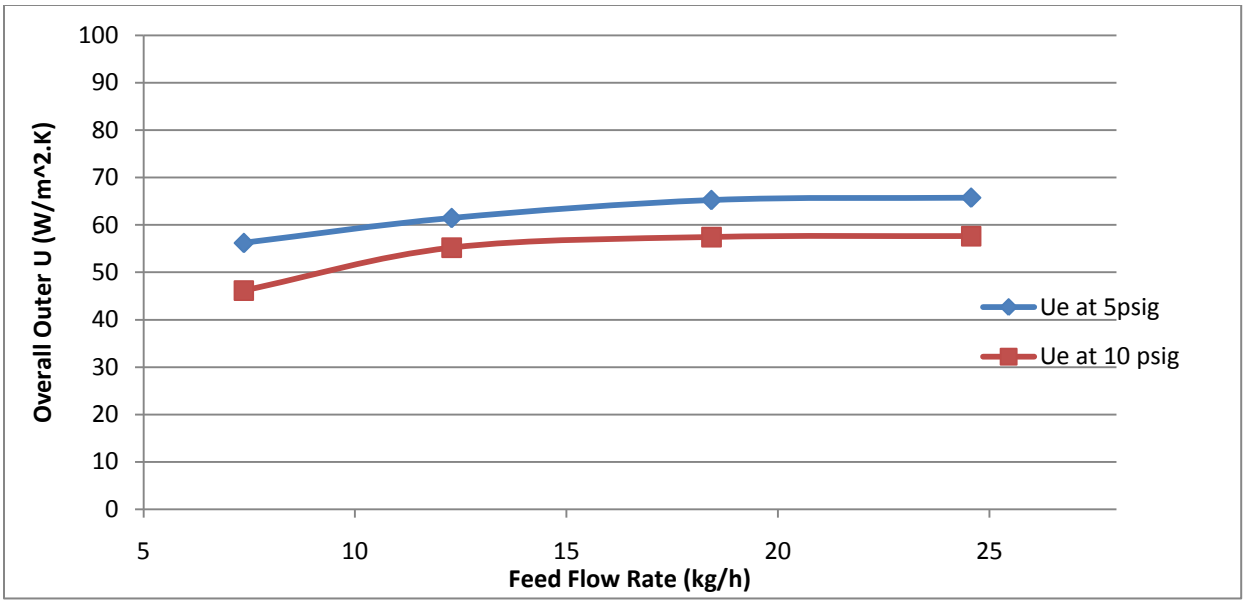


Figure 20: Comparison of Ue at variable steam pressures

The trends for UE presented in this graph could have been easily predicted without going through the calculations. As it has been discussed above that the heat energy lost to the environment QE remains more or less constant it is intuitive to deduce that the outer heat transfer coefficient would also be constant. Heat transfer coefficient is the measure of resistance offered to the flow of heat and if the heat flowing from the system to the atmosphere does not change, it means that the resistance is also not changing. This deduction is verified by the graphical representation of the results above where it can be seen that QE at 5 and 10 psig is a straight line except for the first data point in both cases. As mentioned above the first data point has been a source of anomaly in most of the calculations due to the system not being in steady state when this point was recorded, hence it has affected all the following calculations.

As explained in the case of QE at different pressures, the reason for Ue being slightly higher is the reason that room temperature might have been lower on the day the system was run at 5psig. And since U Is directly proportional to Q, if QE at 5 psig is slightly higher, Ue at 5 psig should also be slightly higher as shown in Figure 25 above,

For the case of the inner overall heat transfer coefficient the evaporation process was broken into two parts. First, where the feed solution gains heat to start boiling was considered the lower limit of the tube which has a separate heat transfer coefficient involving a temperature driving force that raises the feed from its initial temperature to the boiling point of water. The second part is the process of conversion of the boiling solution to water vapor. The tube length for this case was the place where feed starts boiling to the upper end of the tube. This was called the upper part of the process and called the upper heat transfer coefficient which includes the driving force that pushes the temperature of the solution from boiling point of water to the outlet temperature at which the vapor exits the evaporator tube.

The results of the of the lower heat transfer coefficient for the two different steam pressures are presented in Figure 26 below.

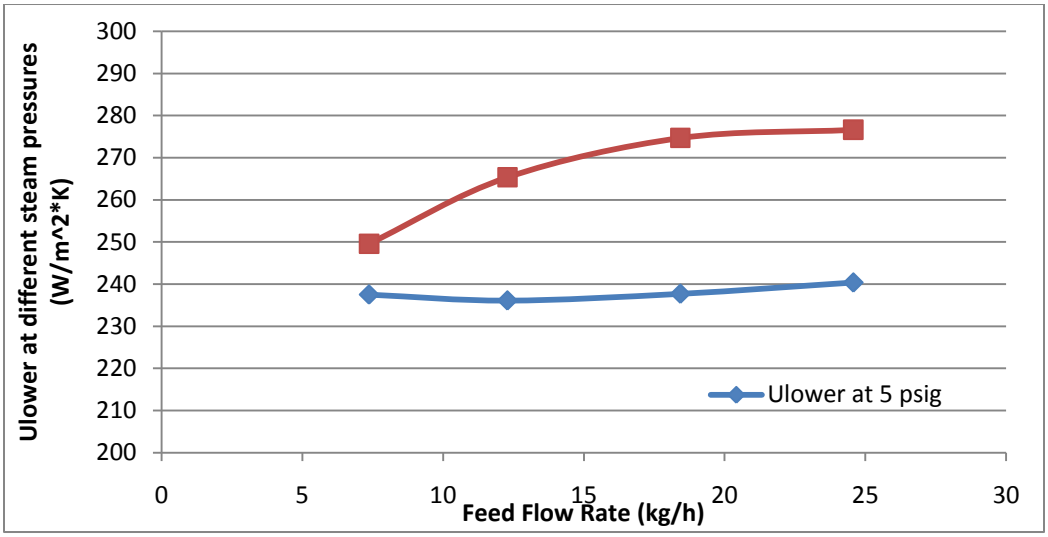


Figure 27: Comparison of the Ulower at different steam pressures

Comparing the general trends of the lower heat transfer coefficient it can be seen that at both pressures, the heat transfer values are more or less constant as the feed flow rate is increases. They are constant around 300 W/m^2.K. This result leads us to conclude that the heat transferred from the steam to the feed in the lower part of the tube (heat supplied for raising the temperature of the feed) is constant irrespective of the steam pressure or the feed flow rate. The slight discrepancy in this graph is that Ulower at 10 psig seems to be higher than Ulower at 5 psig. However, on close inspection one can see that there is not much difference between the values of Ulower at the different pressures and this can be verified from the table below which gives a summary of Ulower values.

Feed Flow Rate (kg/h)	Boiling Height at 5psig (inches)	Ulower ay 5 psig(W/m^2*K)	Boiling Height at 10 psig (inches)	Ulower ay 10 psig (W/m^2*K)
7.371072	16.7	237.528	15.5	249.56
12.28512	28	236.114	24.5	265.347
18.42768	40	237.719	35.5	274.69
24.57024	60	240.407	47	276.638

Table 4: Comparison of boiling heights and Ulower at different pressures and feed flow rates

Evaluating the values of Uupper we find that overall upper heat transfer coefficient is again constant as the feed flow rate is increased. We saw a similar trend in Ulower values as well, where Ulower remained constant as the feed flow rate was increased. These results can be a little confusing as we would expect U lower to increase with increasing feed flow rate because Q lower increases. Similarly we would expect Uupper to decrease with the feed flow rate as because Q upper decreases as the feed flow rate is increased. This behavior of Uuuper and Ulower can be attributed to the affect that the total surface area for heat transfer has on the overall heat transfer coefficient. In case of Ulower, Q lower does increase with feed flow rate but so does the boiling height of the feed. This means that there is

more energy being transferred to the process by the steam which should increase the HT coefficient but this transfer is taking place over a larger surface area thereby decreasing the HT coefficient. This increase and decrease average out the heat transfer coefficient which ends up being constant. Similarly, for Uupper decreases with increasing flow rate which should decrease Uupper, but the area of heat transfer is decreasing too which should increase the U, and both these factors end up averaging out Uupper to be a constant. The trend followed by Uupper at different pressures and feed flow rates is shown in Figure 27 below.

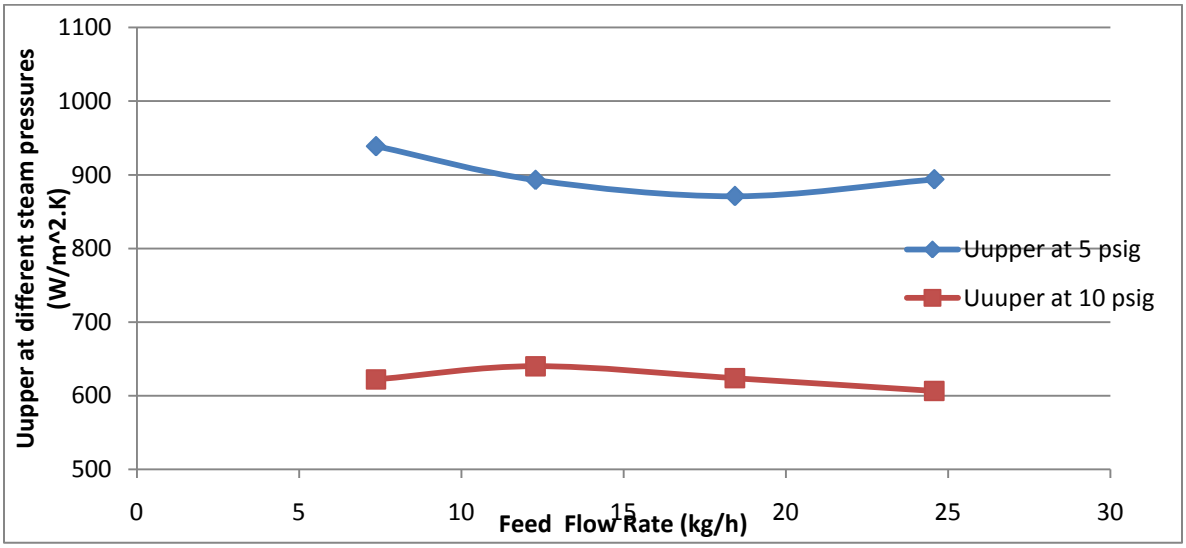


Figure28: Comparison of the U(upper) at different steam pressures

The operating pressure of steam does have an effect on the upper heat transfer coefficient. Since at a higher pressure, a larger amount of water evaporates because of the higher temperature of steam. However, in the upper part of the tube the temperature difference between the steam and the vapor is also higher thereby decreasing the effect of the higher heat transfer. At 5 psig this temperature difference is around 8K whereas at 10 psig this difference is around 16 K and this larger delta T decreases the Uupper at higher pressure of steam. That's why we see that in the graph above, Uuuper is higher at lower pressure and lower at higher pressure.

Comparison of U lower and U upper

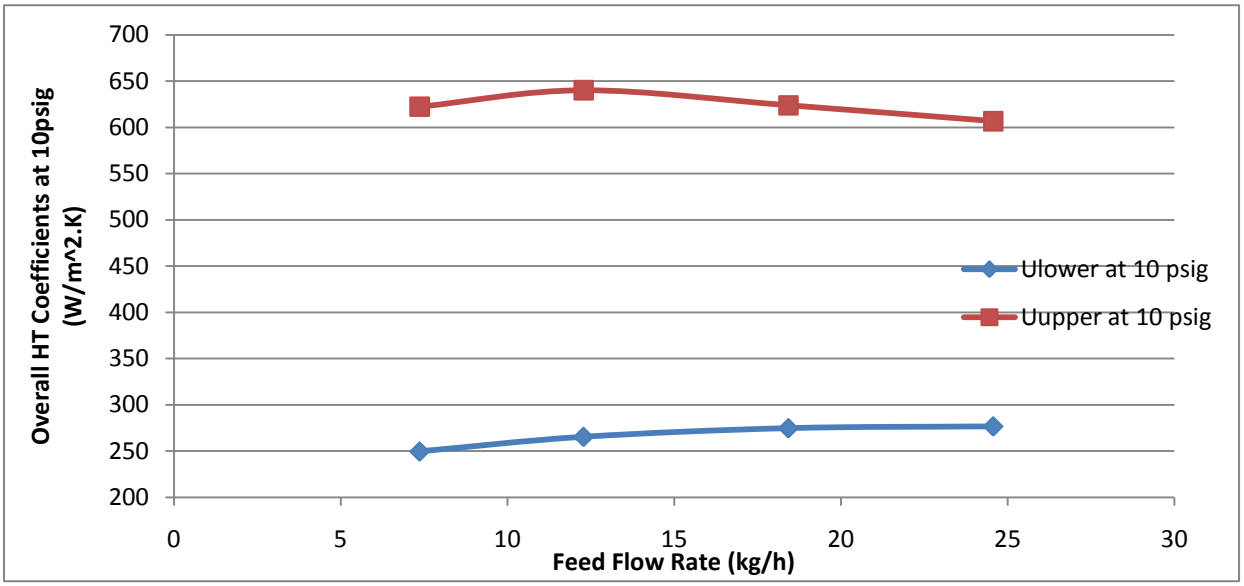


Figure29: Comparison of U lower and U upper at 10 psig

As already stated above both Ulower and Uupper are constant as the feed flow rate increases and the reasons for this trend have been described above. At the same steam pressure and same flow rates, Uupper is larger than Ulower and the reason for this is that a higher amount of heat is transferred in the upper part of the tube as that is where a phase change is taking place and since latent heat of water is very large, U upper is larger than Ulower where only heating of the feed is taking place.

Part 4: Results for Evaporator Economy and Capacity

The success of any process is totally dependent on a cost-benefit analysis. If the cost of the process exceed the profit then the process is considered to be inefficient and uneconomic and if the benefits are higher than the costs the process if considered to be beneficial and efficient. The results of evaporator capacity are presented in Figure 29 below.

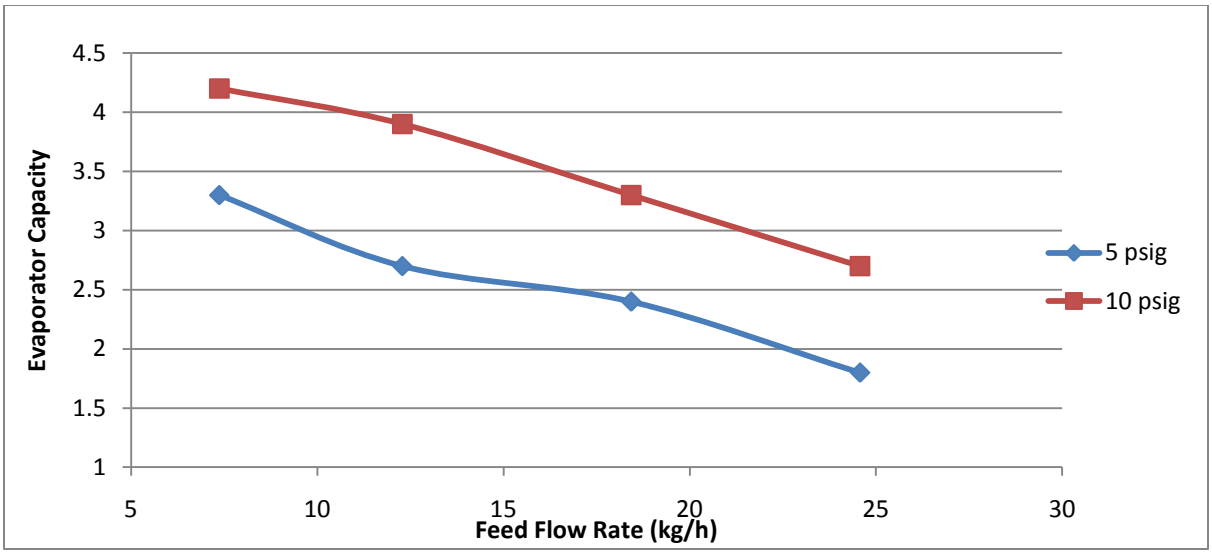


Figure 30: Evaporator Economy at variable steam pressures

The graph above illustrates that as the feed flow rate increases the evaporator capacity decreases which makes sense since capacity is directly the measure of the amount of water evaporated from the feed solution. As discussed earlier the condenser solution flow rate decreases with increased feed flow due to smaller contact time between steam and feed and less energy being transferred to the feed solution. The capacity however increases as the steam pressure is increased (agrees with the results of condenser solution flow rate) as discussed in Section 1 table 1 and 2.

The most important measure of the success of the evaporator is its economy. It was stated in Part 3 earlier that the heat lost to the environment in an evaporator is always higher than the heat gained by the process which makes the evaporator rather inefficient and the only reason to use it for commercial purposes would be if the economy of the evaporator outweighs the energy loss considerations. This can be deduced from the Figure 30 below.

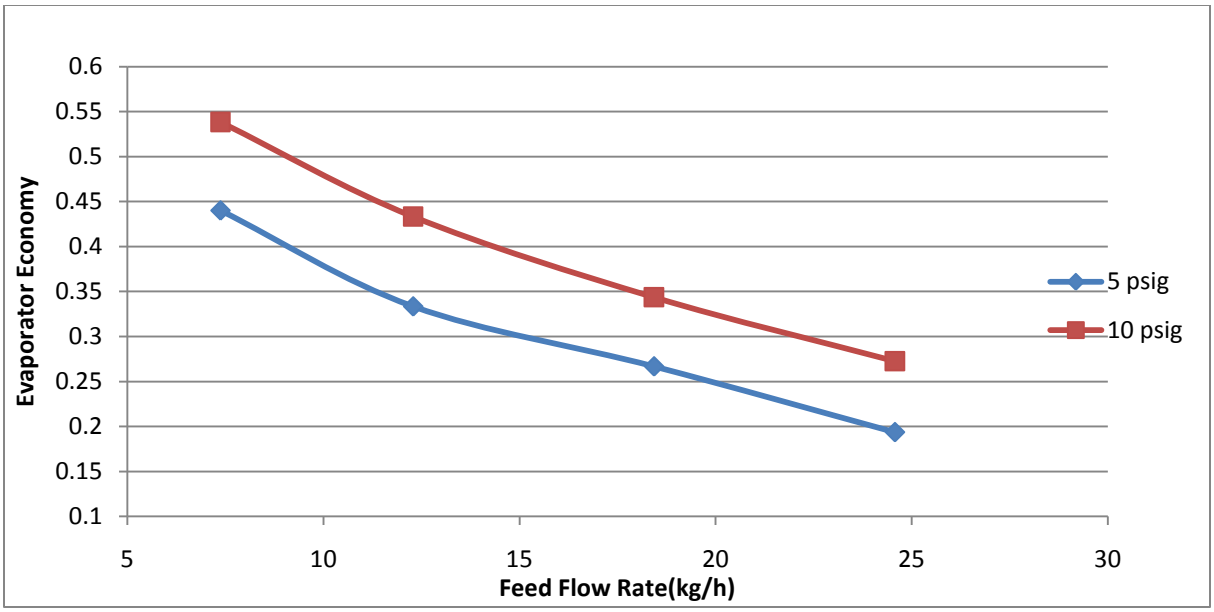


Figure 31: Evaporator Economy at variable steam pressures

The graph above indicates that just like capacity evaporator economy also decreases with an increase in feed flow rate. The economy is higher for a higher steam pressure compared to a lower steam pressure. This can be explained by the fact that at higher flow rates less water is being vaporized and therefore the product solution is less concentrated and at higher steam pressures and constant flow rate more water is vaporized from the feed solution resulting in a more concentrated product.

In conclusion, the way to elicit the best performance from the climbing film evaporator is to operate it at low flow rates and high pressure as this would result in a more concentrated result and better economy. The flip side is that we have to operate the evaporator for longer times since it takes longer for the process to attain steady state at lower flow rates. This will also result in a higher resistance to heat transfer in the form of higher heat transfer coefficients at higher pressures. In the end a balancer has to be maintained between operating time and energy and product specifications and economy.

Part 5: COMSOL Modeling Results

Material balance results

From Figure 31 we can see the concentration of glycerol throughout the length of the climbing film evaporator. The mixture enters at the bottom of the evaporator with a concentration of 10 percent glycerol in water and exits at the top with a concentration of 12.8 percent glycerol in water. In accordance with the behavior seen in the experimental run, we can see from Figure 32 that the concentration of glycerol in the solution starts increasing only once it reaches the boiling temperature of water, which is attained at the boiling height in the tube denoted by the horizontal line in Figure 31. Our experimental data showed an entering mixture composition of 10 percent glycerol in water and an exit composition of 12 percent glycerol in water. Thus, we can conclude that the results from the lab experiment and COMSOL model are comparable.

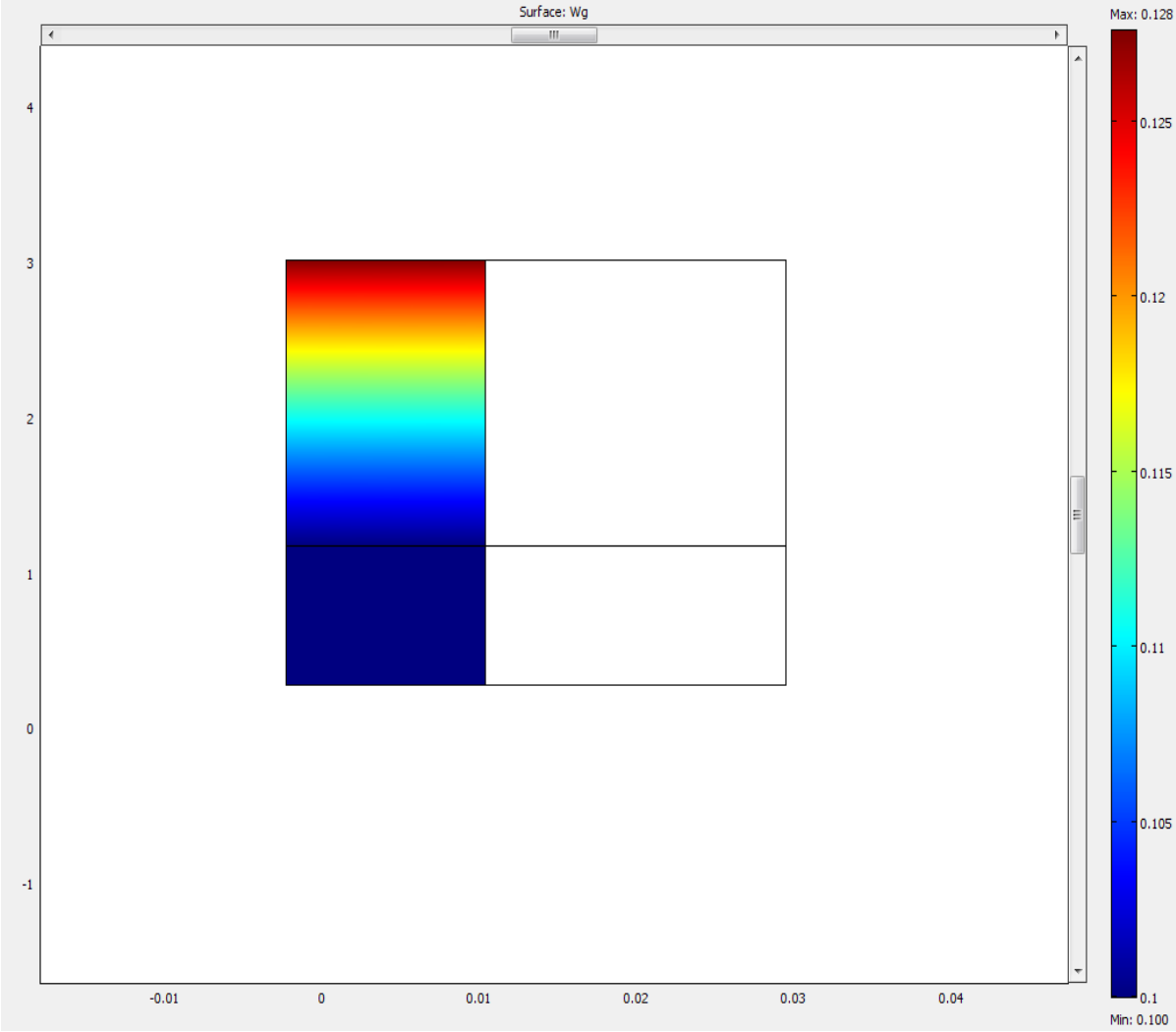


Figure 32: Concentration of glycerol throughout the length of the climbing film evaporator

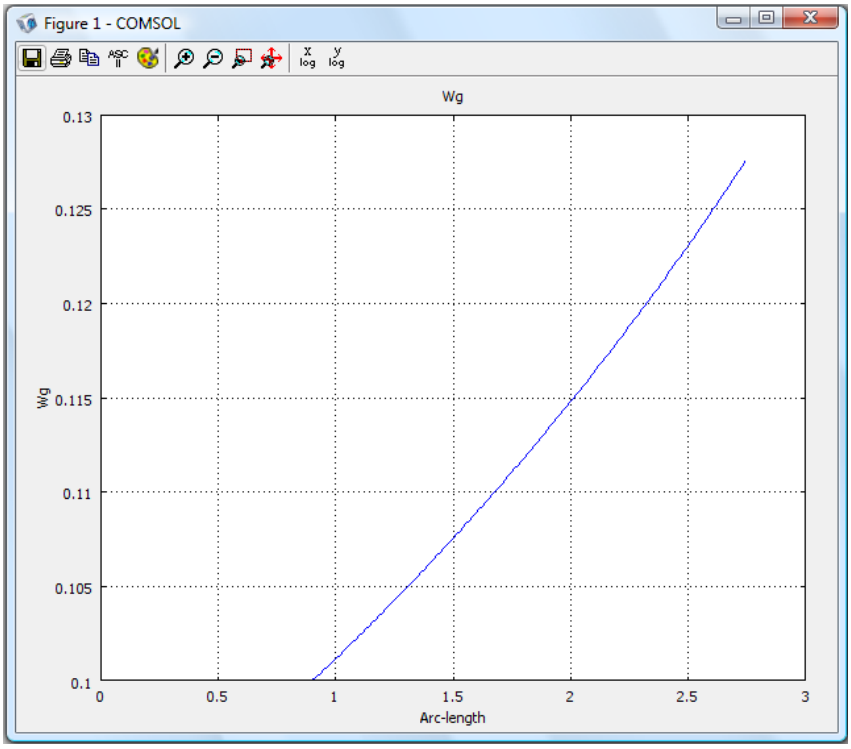


Figure 33: Glycerol in water concentration profile

Energy balance results

Using COMSOL boundary integration we solved for the total heat given by the steam and using formula [7] we solved for the steam trap flow rate. Results are presented in table 5. The graphs prepresented in the main report are the results for a feed flow rate of 300ml/min at a steam pressure of 10 psig. All the other graphs and figures corresponding to these numerical results are given in Appendix C.

Feed Flow Rate = 120 ml/min	COMSOL	Experimental	% Error
Qs	5422.67	4801.34	0.0909
ms	8.579	7.8	0.1145
Wg	0.244	0.17	0.3032

Table 5: Comparison of COMSOL and Experimental Results at Flow Rate 1

Feed Flow Rate = 200 ml/min	COMSOL	Experimental	% Error
Qs	5544.13	5540	0.000075
ms	8.773	9.0	-0.026
Wg	0.16	0.14	0.125

Table 6: Comparison of COMSOL and Experimental Results at Flow Rate

Feed Flow Rate = 300 ml/min	COMSOL	Experimental	% Error
Qs	5775.36	5909.33	-0.0232
ms	9.13	9.6	-0.050
Wg	0.128	0.12	0.0625

Table 7:Comparison of COMSOL and Experimental Results at Flow Rate 3

Feed Flow Rate = 400 ml/min	COMSOL	Experimental	% Error
Qs	5844.44	6094	-0.0427
ms	9.3	9.9	-0.0706
Wg	0.122	0.11	0.0984

Table 8: Comparison of COMSOL and Experimental Results at Flow Rate 4

According to the results the highest % error for concentration of glycerol is for the very first flow rate of 120 ml/min. The % error for this run is around 30%. The COMSOL model predicts the final concentration of around 25% while the experimental result is only 17%. However, the COMSOL model makes sense since in our theoretical calculations; the final concentration of glycerol was found out to be 23% which is close to what the model predicts. The % error for the steam trap flow rate is also the highest for this particular run and so is the heat given off by the steam. All these results re-iterate the explanation given in the previous sections of the report about the first run for both steam pressure being faulty because the system, not having attained steady state when the measurements were made. Another reason for this discrepancy in COMSOL and experimental values for this run and also the other runs was the fact that COMSOL models the experiment ideally whereas in real life there could be a leak from the steam to the environment or the ambient temperature could be higher or lower than what we specified in COMSOL. The percent error in the steam trap flow was only 5% for the 300ml/min run. Thus, we concluded that by finding better heat transfer coefficients to use in the COMSOL model we can simulate the experiment with more accuracy.

As the flow rates of the feed increase we see, that the experimental results become more aligned and in agreement with COMSOL results. This indicates that the latter experimental runs were much better in terms of steady state achievement.

Figure 34 represents the temperature profile of the mixture inside the inner tube with respect to the height of the evaporator. The graph illustrates how the temperature of the mixture inside the inner tube increases exponentially until reaching the boiling height measured at 0.9017m. The temperature then levels off and remains constant at 372.8 Kelvin. This result agrees with the fact that temperature doesn't increase when there is a phase change.

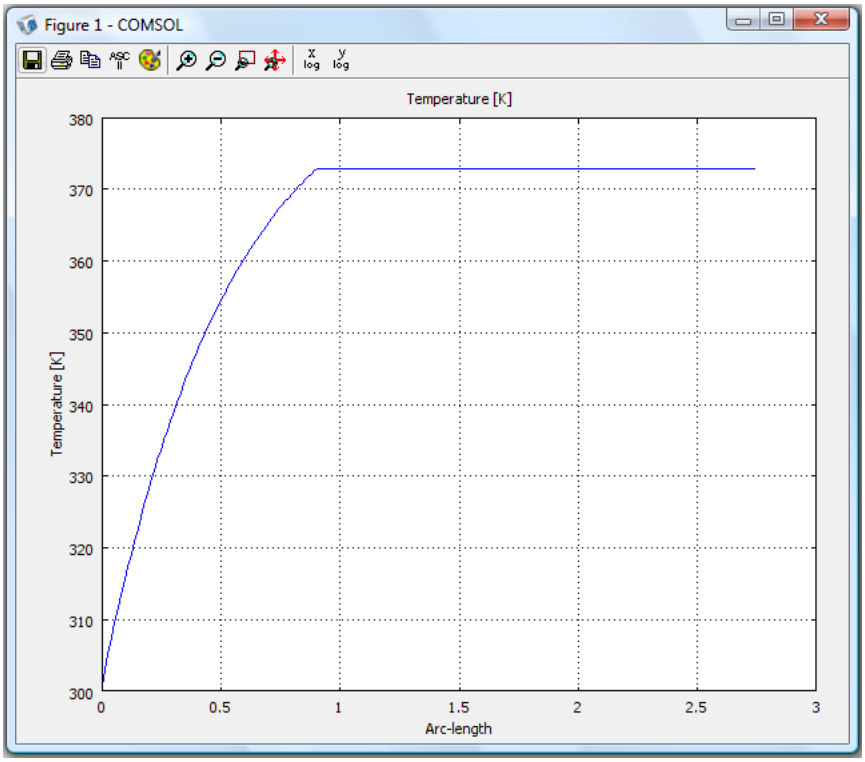


Figure 34: Temperature profile

Comparison of COMSOL Results at Different Pressures

Feed Flow Rate = 300 ml/min	COMSOL	Experimental	% Error
Qs	4499.48	5585	-0.19436
ms	8.11	9.0	-0.10845
Wg	0.112	0.11	0.017857

Table 9: Comparison of COMSOL and Experimental Results at Flow Rate 3 and 5psig steam pressure

Comparing the results for the same feed flow rate at two different steam pressures is indicates that the concentration of glycerol is modeled very well at this flow rate by COMSOL and is close to the experimental value. Similarly, the steam trap flow rate indicated by COMSOL is very close to the flow rate we obtained experimentally. There is a marginal difference between the modeled value for heat lost by steam and the actual value obtained experimentally. As explained previously, the reason for this could be that the ambient temperature might have been different on the day we made the 5 psig run, thereby affecting the flow of heat from the steam to the atmosphere. However, since in COMSOL we indicated the ambient temperature to be the average room temperature, it didn't take into account any fluctuations and again gave us results in an ideal case thereby the difference in the heat loss values.

Heat Transfer Coefficient Results

Figure 6 presents the heat transfer coefficients used in COMSOL and the ones calculated in the experiment.

	COMSOL	Experimental	% Error
Uu at 5 psig	900	871.074	0.03214
Uu at 10 psig	800	623.816	0.22023
Ul at 5 psig	300	237.719	0.207605
Ul at 10 psig	400	274.69	0.313275

Table 10: Experimental and COMSOL Heat Transfer Coefficients

As we can see for the upper heat transfer coefficient there is only 3% at 5 psig difference. In contrast, for the lower heat transfer coefficient there is 20% difference. At 10 psig, U upper has an error of 22% and U lower 31%. The large difference between the COMSOL and the experimental upper heat transfer coefficients might be again because COMSOL models the experiment ideally.

Additionally, in the experiment there is the formation of a liquid film caused by the condensation of water in the steam side that affects significantly the upper heat transfer coefficient causing it to increase. In COMSOL this behavior is not modeled.

Conclusions

Based on our results from the experiment we can conclude that as we increase the operating steam pressure the flow rate of the product stream decreases however the concentration of glycerol in the product stream increases. The concentration of the glycerol in the product decreases with an increase in the feed flow rate. Hence we can safely conclude that in order to get the highest possible product concentration we should run the evaporator on high steam pressure and low feed flow rates. Our results also validated the assumption that the condenser solution is in its totality water with negligible glycerol present. We also concluded that the concentration profile for glycerol in the product obtained experimentally is in accordance with theoretical predictions except for the run at 120 ml/min. This anomaly was attributed to the system not being at steady state when the data was recorded.

From our energy balance results we concluded that the heat gained by the process as well as the heat lost to the environment both increased linearly with an increase in steam pressure and feed flow rate. We concluded that the heat gained by the process shows a linear increase at both steam pressures and is higher for the higher steam pressure. Energy lost to the environment by the steam however remains constant and the explanation for this has been provided in the energy balance results section.

From our heat transfer coefficient results we concluded that the overall outer heat transfer coefficient remains more or less constant. As mentioned above, since the energy lost to the environment remains constant with an increase in feed flow rate and also a constant at different steam pressures, it is intuitive that the outer heat transfer coefficient which is a measure of the rate of heat transfer to the atmosphere should remain constant

Both the inner heat transfer coefficients show a slight increase with an increase in the feed flow rate although they can be considered constant for all practical purposes and this is what was done to model the process in COMSOL We also concluded that the inner heat transfer coefficients are very sensitive to the height at which the phase change in the feed occurs. Even a change of five inches in this height can change the coefficients by a large magnitude.

From the evaporator performance results we can conclude that the evaporator capacity decreases with an increase in the feed flow rate due to the reduced contact time between steam and the feed solution. The capacity increases with increasing steam pressure. The most important measure of the performance of the evaporator is the evaporator economy. We concluded that the evaporator economy decreases with an increase in feed flow rate and increases with an increasing operating steam pressure. This conclusion validates our first assumption that in order to get the best results we should run the evaporator at lower feed flow rates and higher operating steam pressures.

For the feed rate of 300 ml/min, COMSOL approximated very well predicting a product concentration of 12.8 percent glycerol in water whereas in the experiment the measured concentration was 12 percent. The energy balance results predicted were also close to the experimental values obtained. COMSOL reported a value of 5775 W of heat given off by the steam and in the experiment the calculated heat given by the steam was 5909 W. The slight discrepancy in values was attributed to the fact that COMSOL models the experiment ideally whereas in the actual experiment there could have been factors such as a leak from the steam to the environment that could have affected the calculations. The temperature profile of the feed generated by COMSOL matched well with the profile predicted in our calculations. There was a good agreement between the upper heat transfer coefficients calculated

in the experiment and COMSOL with only 3% percent difference however the highest error was 30%. We believe this difference is due to the fact that in the experiment a thin film liquid forms in the glass wall caused by the condensation of water in the steam side that affects significantly the upper heat transfer coefficient causing it to increase. In COMSOL this behavior is not modeled.

Based on our COMSOL results and the experimental data, we can conclude that COMSOL is a very effective way for simulating a climbing film evaporator given the correct heat transfer coefficients, heat flux expressions, boundary conditions, and concentrations.

Recommendations

The accuracy of the data obtained from this experiment can be increased by implementing new process conditions or changing some preexisting ones. First, more time should be allowed for each run to attain steady state. We waited 20 minutes for each run before recording the data. Although, this time might have been enough for some runs it might have been insufficient for others. We recommend future experimenters to measure the concentration of glycerol in the product every 10 minutes and once the concentration stops changing steady state would have been attained for sure. We also recommend doing additional runs using more feed flow rates and steam pressures in order to verify the trends shown by data.

We also recommend that certain variables such as ambient temperature (especially if the runs are conducted on different days) should be taken into account for every single flow rate. As explained in parts of the report the ambient temperature value might have affected some of the heat balance calculations and been a source of error. In order to eliminate the possibility of this error, all temperature should be recorded for each run.

The model that was generated on COMSOL was a basic simulation of the evaporator process. In reality the climbing film evaporator is a complex unit operation. For simplicity in our model we specified the height at which the feed boils instead of asking COMSOL to model it. We also provided the upper and lower heat transfer coefficients instead of letting the model calculate these according from the process conditions that we input. For future models we suggest to find ways of making COMSOL calculate the heat transfer coefficients and the boiling height.

References

[1] McCabe, W. L., Smith, J. C. & Harriott, P. (2005). Unit Operations of Chemical Engineering (7th edition). In (E. D. Glandt, M. T. Klein & T. F. Edgar, Eds.). USA: Mc-Graw Hill.

[2] Minton, Paul.E. (1986). Handbook of Evaporation Technology. (1st edition). Noyes Publication. Park Ridge, New Jersey.

[3] Ribiero, C.P & Andrade Cano M.H.(2001). A heat transfer model for the steady-state simulation of climbing –falling –film plate evaporators. Journal of Food Engineering. Pages 309-320.

[4] Pacheco, C.R.F & Frioni L.S.M. (2001=3). Experimental results for evaporation of sucrose solution using a climbing film evaporator. Journal of Food Engineering. Pages 471-480.

[5] Young, B.R & Allen, R.M (1995) Multi-Input, Multi Output Identification of a Pilot- Plant Climbing Film Evaporator. Control Engineering Practice. Pages 1067-1-73

[6] Glover, William B. (2004). Chemical Engineering Progress- Selecting Evaporators for Process Applications. AIChE Journal. Pages 26-33.

[7] Sheikholesami, Roya., Yu, Hong, Doherty, William.(2004).Flow Boiling Heat Transfer of Water and Sugar Solutions in an Annulus. AIChE Journal Vol 50, No. 6. Pages 1119-1-1128.

[8] Haiyuan Biological. Rising thin film evaporator. Retrieved on April 24, 2010, from http://www.wxhysh.com/products/prod12.htm

[9] Author Stream. Evaporation. Retrieved on April 27, 2010, from http://www.authorstream.com/Presentation/jitpatel21-352080-evaporation-science-technology-ppt-powerpoint/

[10] Genemco. Aloe Vera Processing Equipment- Rising Film Evaporator. Retrieved on April 28, 2010, from http://www.genemco.com/aloe/rising.html

[11] Filter Machinec Manufacturers. Water Distillers. Retrieved on April 28, 2010, from http://www.filters-machines.com/water-distillers.html

[12] De Dietrich Process Systems. Evaporation-Climbing Film and Natural Circulation Evaporator. Retrieved on April 27, 2010, from http://www.ddpsinc.com/TeachingUnits/qts4.html

[13] COMSOL Multiphysics. Chemical Engineering Module-Water Treatment. Retrieved on April 25, 2010, from http://www.comsol.com/products/chem/

Appendices

Appendix A- Sample Calculations

For all the sample calculations given below data recorded for the third run on day 2 of the experiment (Feed flow rate of = 300 ml/min at steam pressure of 15 psig) was used.

Sample Calculations for Mass Balance

To calculate the mass balance we first need the mass flow rate of the feed solution. Since we measured the flow rate of the feed using a flow meter, it is in the units of ml/min and we need to convert it to kg/h. The first step to achieve this is to find the density of the feed solution. Since the solution is a mixture of water and glycerol, the density of this solution will be a combination of the densities of water and glycerol. The calculations are as follows:

$$Density\ of\ Glycerol\ at\ 25^oC\ (\rho_{Gly})\ = 1.261 \frac{g}{cm^3}$$

$$\%\ of\ Glycerol\ in\ the\ feed\ solution\ (x_f) => \ 10\% = 0.1$$

$$Density\ of\ Water\ at\ 25^oC\ (\rho_w)\ = 0.9974 \frac{g}{cm^3}$$

$$\%\ of\ Water\ in\ the\ solution\ (x_{f_w}) => (\ 1 - 0.1) = 0.9$$

$$Density\ of\ the\ solution\ (\rho_{feed}) = (\rho_{Gly} * x_{f_{gly}}) + (\rho_w * x_{f_w})$$

$$Density\ of\ the\ solution\ (\rho_{feed}) = (1.261 * 0.1) + (0.9974 * 0.9) = \mathbf{1.02376}\ \frac{g}{cm^3}$$

Calculating the Mass Flow Rate of the Feed:

$$Volumetric\ Flow\ Rate\ of\ the\ Feed\ (\dot{V}) = 300 \frac{ml}{min}$$

$$Density\ of\ the\ solution\ (\rho_{feed}) = 1.02376\ \frac{g}{cm^3}$$

$$Mass\ Flow\ Rate\ of\ the\ Feed\ (\dot{m}_f) = \dot{V} * \rho_{feed}$$

$$Mass\ Flow\ Rate\ of\ the\ Feed\ (\dot{m}_f) = 300\ \frac{ml}{min} * 1.02376 \frac{g}{cm^3} * \frac{cm^3}{ml} = 307.128 \frac{g}{min}$$

$$Mass\ Flow\ Rate = 307.128 \frac{g}{min} * \frac{60\ min * kg}{1000\ g * h} = 18.43 \frac{kg}{h}$$

$$Mass\ of\ glycerol\ in\ the\ feed\ (m_{f_{gly}}) = 0.10 * 18.43 \frac{kg}{h} = 1.843 \frac{kg}{h}$$

$$Mass\ of\ water\ in\ the\ feed (m_{f_w}) = 0.90 * 18.43 = 16.585 \frac{kg}{h}$$

Calculating the Mass Flow Rate of the Condenser Solution (Water):

$$Measured\ Mass\ Flow\ Rate\ of\ the\ condensate(\dot{m}_v) = 55\frac{g}{min}$$

$$Mass\ Flow\ Rate\ of\ the\ condensate(\dot{m}_v) = 55\frac{g}{min} * \frac{60\ min * kg}{1000\ g * h} = 3.3\frac{kg}{h}$$

Calculating the Mass Flow Rate of the Concentrated Glycerol Solution (Product): (with 12 % glycerol concentration)

$$Mass\ flor\ rate\ of\ the\ product\ solution(\dot{m}_P) = \dot{m}_f - \dot{m}_v$$

$$Mass\ flor\ rate\ of\ the\ product\ solution = 18.43\frac{kg}{h} - 3.3\frac{kg}{h} = 15.1277\frac{kg}{h}$$

$$Mass\ \ of\ glycerol\ in\ the\ product\ solution\ \left(m_{p_{gly}}\right) = 0.12 * 15.1277\frac{kg}{h} = 1.815\frac{kg}{h}$$

$$Mass\ of\ water\ in\ the\ product\ solution\left(m_{p_w}\right) = 15.1277 - 1.815 = 13.313\frac{kg}{h}$$

Going back to the calculations for the condenser solution we can now calculate the mass of water and glycerol in the solution that condenses out of the evaporator in the product solution

$$Mass\ of\ glycerol\ in\ the\ condensate\ \left(m_{gly}\right) = m_{f_{gly}} - m_{v_{gly}}$$

$$Mass\ of\ glycerol\ in\ the\ condenser\ solution\ \left(m_{gly}\right) = 0\ \frac{kg}{h}$$

*<u>**Assumption-**</u> We assume that all the liquid that comes out of the evaporator is water since the temperature and pressure of the system are high enough to boil and evaporate just water (temp was 1030C and pressure was 15 psig) and glycerol has a boiling point of 290oC and thus does not evaporate at all and most of it goes directly into the product solution as a liquid.

$$Mass\ of\ water\ in\ the\ condenser\ solution\ (m_{water}) = m_{f_{gly}} - m_{v_{gly}}$$

$$Mass\ of\ water\ in\ the\ condenser\ solution\ (m_{water}) = 16.6\frac{kg}{h} - \frac{kg}{h} = 13.313\frac{kg}{h} =$$

$$\%\ of\ water\ in\ the\ condenser\ solution = \frac{3.273\frac{kg}{h}}{3.3\frac{kg}{h}} * 100 = 99.2\%$$

The above calculation supports the assumption made above that in these kinds of experiments the solution that condenses out from the evaporator is just water and we can see there is only a minimal quantity of glycerol (0.4%) in the condenser solution. This is an intuitive assumption since we are only heating the feed solution to the boiling point of water and hence water is the only liquid in the feed

solution that actually evaporates. The little quantity of glycerol that is present in this solution could be attributed to the fact that maybe there was some glycerol already present in the tube or the bucket that was used to collect the solution or that perhaps some glycerol did slosh over into the evaporator along with water vapor.

Checking the Mass Balances:

Condenser Solution+ Concentrated Glycerol Solution= Feed Solution

Overall Mass Balance:

$$3.3\frac{kg}{h} + 15.127\frac{kg}{h} = \mathbf{18.427}\frac{\boldsymbol{kg}}{\boldsymbol{h}}$$

$$Actual\ feed\ flow\ rate = 18.427\frac{kg}{h}$$

Glycerol Mass Balance:

$$0\frac{kg}{h} + 1.815\frac{kg}{h} = \mathbf{1.8153}\frac{\boldsymbol{kg}}{\boldsymbol{h}}$$

$$Actual\ glycerol\ flow\ rate\ in\ the\ feed = 1.843\frac{kg}{h}$$

$Error\ Calculation$= (1.842-1.815)/1.845=2.6%

Water Mass Balance:

$$3.273\frac{kg}{h} + 13.313\frac{kg}{h} = \mathbf{16.586}\frac{\boldsymbol{kg}}{\boldsymbol{h}}$$

$$Actual\ water\ flow\ rate\ in\ the\ feed = 16.585\frac{kg}{h}$$

$Error\ Calculation$= (16.585 -16.586)/ 16.585 =-0.006%

The slight errors in the mass balance checks could have been due to errors made in collecting the condenser solution from the evaporator. We could have made mistakes in timing our collection interval and could have spilled some of the condensate from the bucket in the process of weighing the bucket to determine the flow rate. As stated earlier some glycerol could have possibly sloshed over into the evaporator along with water vapor giving us an error margin in the mass balance for glycerol and also giving us a larger value for water collected than that was in the feed (since mass cannot be created or destroyed and the overall mass balance does agree then this is the most likely explanation for the discrepancy).

Another source of error might be uncertainties in the instruments being used to measure different quantities such as time, flow readings from the pump or the specific density probe.

The final and possibly the most prominent source of error could have been the fact that the process had not yet attained steady state when we made the measurements. If this did happen indeed, then it could have altered our data and given us faulty measurements.

Theoretical Calculations for % glycerol in the product solution

$$\dot{m}_f * x_{f_{gly}} = \dot{m}_p * x_{p_{gly}}$$

$$\%\ Gly\ in\ the\ product\ solution\ (x_{p_{gly}}) = \frac{\dot{m}_f * x_{f_{gly}}}{\dot{m}_p}$$

$$\%\ Gly\ in\ the\ product\ solution\ \left(x_{p_{gly}}\right) = \frac{18.427 * 0.1}{15.1277} = 12.5\%$$

The theoretical % of glycerol in the product solution is slightly higher than the % glycerol actually measured in the product solution. This slight discrepancy can be attributed to experimental error. We might have made mistakes in cooling down the product solution to 25^{o}C *(we might have over or under cooled) which might have given us faulty readings. We also could have made a mistake in reading off the % of glycerol from the percentage-specific gravity charts. The highest deflection from theoretical data is the very first run conducted at 120 ml/min of feed flow rate. The possible explanation for this error is that the system had not yet attained steady state when the measurements were made thereby skewing the data by such a huge margin for that data point. All the other data points are within $\pm$ 1% of the theoretical prediction.

Sample Calculations for Energy Balance

Heat Flow from Steam to Water (Q)

$$Q_S = \dot{m}_s \lambda_s$$

$$Q_S = Q_p + Q_E$$

$$Q_S = \dot{m}_v \lambda + \dot{m}_f C_p \Delta T + Q_E$$

Where:
$Q_E = Energy\ losses\ to\ the\ environment$
$\dot{m}_v Flow\ rate\ of\ the\ condenser\ solution$
$\lambda = Latent\ Heat\ Capacity\ of\ the\ solution$

$\dot{m}_f = Mass\ flow\ rate\ of\ the\ feed\ solution$
$C_p = Specific\ Heat\ Capacity\ of\ the\ feed\ solution$
$\Delta T = \left(T_{boiling} - T_{feed}\right)$

The heat flowing from the steam to the feed solution (Q_P heat gained by the process) is expressed in watts and calculated from an energy balance on the feed and the outgoing streams which are:

- Product(saturated hot liquid phase liquid leaving the evaporator)
- Condensate (saturated vapor leaving the evaporator, then being condensed and cooled)

To calculate the energy balance we first need to calculate the latent heat for the solution which comes out of the condenser.(Cite paper here). Latent heat calculations for solution with water as a solvent are not as straightforward as taking an average of the latent heat capacities of the components of the solution. From a paper we are going to us 1600 kj/kg as the heat capacity of the feed solution. We first need to calculate the percentage of water and glycerol in the solution. The calculations are given below:

$$Latent\ Heat\ of\ the\ solution\ (\lambda_{soln}) = \mathbf{1600}\frac{kJ}{kg}$$

Converting the latent heat of the solution from kJ/kg to J/kg

$$Latent\ Heat\ of\ the\ solution\ (\lambda_{soln}) = \mathbf{1600000}\frac{J}{kg}$$

Step 2: Calculating the specific heat capacity of the concentrated glycerol solution:

$$Mass\ flow\ rate\ of\ \ the\ feed\ solution\ \left(\dot{m}_f\right) =\ 18.427\ \frac{kg}{h}$$

$$\%\ of\ Glycerol\ in\ the\ feed\ solution\ \left(x_{f_{gly}}\right) = 10\% = 0.10$$

$$Specific\ Heat\ Capacity\ of\ \ glycerol = 221.9\frac{J}{mol * K}$$

Converting the specific heat capacity of glycerol from J/mol.K to J/kg.K

$$Molecular\ Mass\ of\ Glycerol = 92.09\ \frac{g}{mol}$$

$$Specific\ Heat\ Capacity\ of\ \ glycerol = \frac{221.9}{92.09}\ \frac{J}{gm * K}$$

$$Specific\ Heat\ Capacity\ of\ \ glycerol = 2.409\ \frac{J}{gm * K}$$

$$Specific\ Heat\ Capacity\ of\ \ glycerol = 2.409 * 1000\ \frac{J}{kg * K}$$

$$Specific\ Heat\ Capacity\ of\ \ glycerol = 2409.499\ \frac{J}{kg * K}$$

$$\%\ of water\ in\ the\ feed\ solution\ \left(x_{f_{gly}}\right) = 90\% = 0.90$$

$$Specific\ Heat\ Capacity\ of\ water = 4186\ \frac{J}{kg * K}$$

$$Specific\ Heat\ Capacity\ of\ product\ solution = \left(0.90 * 4186\ \frac{J}{kg * K}\right) + \left(0.10 * 2409.499\ \frac{J}{kg * K}\right)$$

$$Specific\ Heat\ Capacity\ of\ the\ feed\ solution(C_{p_f}) = \mathbf{4008.35}\frac{\boldsymbol{J}}{\boldsymbol{kg * K}}$$

Step 3: Calculating the temperature difference

$$Boiling\ temperature\ for\ the\ feed\ solution(T_b) = 373.15\ K$$

$$Inlet\ temperature\ for\ the\ feed\ solution\left(T_f\right) = 84\ F$$

$$Inlet\ temperature\ for\ the\ feed\ solution\left(T_f\right) = 302.04\ K$$

$$Temperature\ difference\ \left(T_b - T_f\right) = 373.15\ K - 302.04\ K = 71.94K$$

Step 4: Calculating the heat gained by the process:

$$Q_p = [\dot{m}_f C_{p_f} * \left(T_b - T_f\right)] + [\dot{m}_v \lambda_f]$$

Q to boil

Q to evaporate

$$Q_p = (18.427\frac{kg}{h} * 4008.35\frac{J}{kg * K}\ * (71.94K) + \left(3.3\frac{kg}{h} * 1600000\frac{J}{kg}\right)$$

$$Q_p = 10988523\ \frac{J}{h}$$

Converting this from Joules/ hour to Watts:

$$Q_p = \frac{10988523\ \frac{J}{h}}{3600\ \frac{sec}{h}} = 3052.368\ Watts$$

$$Q_p = \mathbf{3.052\ kW}$$

Step 5: Calculating the heat lost by the steam:

$$Q_S = \dot{m}_s \lambda_s$$

Where:

$\dot{m}_s$ is the condensate mass flow rate

λ_s is the latent heat of vaporization for steam

$$Condensate\ Mass\ Flow\ Rate\ (\dot{m}_s) = 9.6\frac{kg}{h}$$

$$Condensate\ Mass\ Flow\ Rate\ (\dot{m}_s) = 0.002667\frac{kg}{sec}$$

$$Q_S = 0.002667\frac{kg}{sec} * 2216000\frac{J}{kg}$$

$$Q_S = 5910\ Watts$$

Calculating the heat lost to the environment using the overall energy balance for the evaporator:

$$Q_S = Q_p + Q_E$$

$$Q_E = Q_S - Q_p$$

$$Q_E = 5910\ W - 3052.368W = 2856.97\ Watts$$

$$Q_E = 2.857\ kW$$

Step 1: Calculating the quality of the steam

$$Enthalpy\ of\ Superheated\ Steam\ (H_{super}) = 1157.42$$

$$Enthalpy\ of\ Saturated\ Steam\ (H_{sat}) = 1150.5$$

$$Enthalpy\ of\ the\ condensate\ (H_{cond}) = 180.16$$

$$Quality\ odf\ Steam\ (X) = \frac{H_{cond} - H_{super}}{H_{cond} - H_{sat}}$$

$$Quality\ odf\ Steam\ (X) = \frac{180.16 - 1157.42}{180.16 - 1150.5} = 1.000334$$

<u>Sample Calculations for Heat Transfer Coefficients</u>

$$\frac{1}{U_A} = \frac{1}{h_i * A_i} + \frac{q}{k_q * A} + \frac{1}{h_o * A_o}$$

For the process of concentrating glycerol using the climbing film evaporator there is an outer overall heat transfer coefficient which is given by the expression:

$$Q_E = U_{O_E} * A_O * \Delta T_{LM}$$

Where:

U_{O_E} is the outer heat transfer coefficient

A_O is the outer surface area of the evaporator

ΔT_{LM} is the Log mean temperature difference or the driving force with steam on the inside and air outside(which in this case is just $(T_s - T_a)$

Since we already calculated the energy lost to the environment we can rearrange the equation to

$$U_{O_E} = \frac{Q_E}{A_O * (T_s - T_a)}$$

$$Heat\ energy\ los\ to\ the\ environment(Q_E) = 2856.97\ Watts$$

$$Temperature\ \ of\ the\ steam\ supplied\ at\ 10\ psig\ (T_s) = 389.15\ K$$

$$Temperature\ \ of\ the\ air\ outiside\ the\ evaporator\ (T_a) = 298.15\ K$$

$$Temperature\ difference\ (T_s - T_a) = 389.15\ K - 298.15\ K = 91\ K$$

Step 1: Calculating the outer surface area of the evaporator:

$$Diameter\ of\ the\ inner\ tube\ (\ D_i) = 1\ in = 0.0254m$$

$$Diameter\ of\ the\ outer\ tube\ (\ D_o) = 2.5\ in = 0.0635m$$

$$Lenght\ of\ the\ evaporator\ (L) = 9\ ft = 2.74m$$

$$Outer\ Surface\ \ Area\ of\ the\ evaporator\ tube\ (A_O) = \pi * D_o * L$$

$$Outer\ Surface\ \ Area\ of\ the\ evaporator\ tube\ (A_O) = 0.5467m^2$$

Step 1: Calculating the overall outer HT coefficient:

$$U_{O_E} = \frac{2856.97\ Watts}{0.5467\ m^2 * 91\ K} = \mathbf{57.4}\frac{\boldsymbol{W}}{\boldsymbol{m^2 * K}}$$

Calculating the inner heat transfer coefficient

In case of the evaporator process the inner heat transfer can be calculated in two ways:

- Consider the entire tube as a whole and neglect the phase change happening in the tube and consider the driving force to be between the feed inlet temperature and the outlet temperature
- Break down the tube into two processes
    - One where the feed gets heated and the height of the evaporator is taken to be the height at which the glycerol solution starts boiling. The heat transfer coefficient using these conditions is called the lower overall heat transfer coefficient.

- And the rest of the process where the water starts to evaporate (the height is considered to be the total height minus the boiling height) is used to calculate the upper heat transfer coefficient.

$$Q_P = U_l * A_l * \Delta T_{LM} + U_u * A_u * \Delta T_{LM}$$

Using the second method:

Step 1: Calculating the lower overall outer HT coefficient:

$$Diameter\ of\ the\ inner\ tube\ (D_i) = 1\ in = 0.0254m$$

$$Length\ where\ the\ solution\ starts\ boiling(L_l) = 35.5\ in = 0.9017m$$

$$Surface\ Area\ of\ the\ lower\ tube\ (A_l) = \pi * L_l * D_l$$

$$Surface\ Area\ of\ the\ lower\ tube\ (A_l) = 0.07195\ m^2$$

$$Q_l = \dot{m}_f * C_p * (T_b - T_f)$$

$$Temeperature\ difference\ (T_b - T_f) = 75.22K$$

$$Q_l = 18.427\frac{kg}{h} * 4180\frac{J}{kg * K} * 75.22\ K$$

$$Q_l = 5794149.8\ \frac{J}{h}$$

$$Q_l = \frac{5794149.8\ \frac{J}{h}}{3600\ sec/h}$$

$$Q_l = 1609.5\ Watts$$

Calculating the HT coefficient

$$Q_l = U_l * A_l * \Delta T_{LM}$$

$$\Delta T_{LM} = \frac{[(T_s - T_f) - (T_s - T_b)]}{\ln\left(\frac{(T_s - T_f)}{T_s - T_b}\right)}$$

$$\Delta T_{LM} = \frac{[(389.15 - 300.92) - (389.15 - 376.15)]}{\ln\left(\frac{(389.15 - 300.92)}{389.15 - 376.15}\right)}$$

$$\boldsymbol{\Delta T_{LM} = 81.434\ K}$$

$$U_l = \frac{Q_l}{A_l * \Delta T_{LM}}$$

$$U_l = \frac{1609.5\ Watts}{0.07195\ \ m^2 * 81.434\ K}$$

$$U_l = 275 \frac{Watts}{m^2 * K}$$

Step 2: Calculating the upper overall outer HT coefficient:

$$Diameter\ of\ the\ inner\ tube\ (\ D_i) = 1\ in = 0.0254m$$

$$Length\ where\ the\ solution\ starts\ boiling(\ L_l) = 72.5\ in = 1.8415m$$

$$Surface\ Area\ \ of\ the\ lower\ tube\ (\ A_l) = \pi *\ D_l *\ L_l$$

$$Surface\ Area\ \ of\ the\ lower\ tube\ (\ A_l) = 0.146954\ m^2$$

$$Q_u = \dot{m}_v * \lambda$$

$$Q_u = 3.3 \frac{kg}{h} * 1600000 \frac{J}{kg}$$

$$Q_u = 1466.67\ Watts$$

Calculating the HT coefficient

$$Q_u = U_u * A_u * \Delta T$$

$$\boldsymbol{\Delta T = 16\ \ K}$$

$$U_u = \frac{Q_u}{A_u * \Delta T}$$

$$U_l = \frac{1466.67\ Watts}{0.146954\ \ m^2 * 16\ K}$$

$$U_l = 624 \frac{Watts}{m^2 * K}$$

Sample Calculations for Capacity and Economy

To calculate the evaporator capacity the relation used is

$$Capacity = \dot{m}_v$$

$$Capacity = 3.3\ \frac{kg}{h}$$

To calculate the evaporator economy the relation used is:

$$Economy = \frac{Capacity}{\dot{m}_{steam}}$$

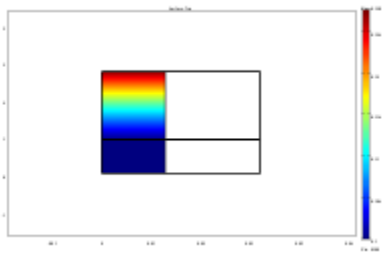
$$Economy = \frac{3.3\ \frac{kg}{h}}{9.6\ \frac{kg}{h}} = 0.34375$$

$$Economy = 34.38\%$$

This means that for every kg/h of steam supplied to the 35 % of water is evaporated and glycerol is concentrated in the product stream.

As the feed flow rate increases at a constant steam pressure the economy decreases and the steam pressure is increased while the feed flow rate is kept constant, both the economy and capacity of the evaporator increase.

Appendix B- COMSOL Model Report



1. Table of Contents

2. Model Properties

Property	Value
Model name	
Author	
Company	
Department	
Reference	
URL	
Saved date	Apr 24, 2010 12:31:56 PM
Creation date	Apr 23, 2010 3:36:33 PM
COMSOL version	COMSOL 3.5.0.603

File name: R:\comsol model v3.mph

Application modes and modules used in this model:

- Geom1 (Axial symmetry (2D))
  - Convection and Diffusion (Chemical Engineering Module)
  - Convection and Conduction (Chemical Engineering Module)
  - Convection and Conduction (Chemical Engineering Module)

3. Constants

Name	Expression	Value	Description
D	1e-9		Diffusivity
Dg	1e-5		Gas diffusivity
K	.6		thermal conductivity
Cp	4180		Heat capacity
rho	1020		Density
Ue	50		Heat transfer coefficient to atmosphere
Ul	500		Lower heat transfer coefficient
Uu	800		Upper heat transfer coefficient
Tf	300.9		Initial temperature
Ts	116+273.15		Steam temperature
Fr	300		Flow rate
Ai	pi*(2.54/200)^2		inner area
vin	Fr/(60*100^3)/Ai		Velocity in
vst	100		Velocity steam
Ta	298		Ambient temperature
Cw0	rho*.9*1000/18		Initial concentration of water

Cg0	rho*.1*1000/92		Initial concentration of glycerol
Cwv0	0		Initial concentration of water vapor
Tb	101+273.15		Boiling temperature
Lt	2.74		Height
Ll	35.5*2.54/100		Boiling height
Lu	Lt-Ll		Upper height
r1	.0127		Inner tube radius
r2	(1.25-.5)*2.54/100+.0127		Outer tube radius
a	2/r1		surfarea1/vol1
a2	2*r2/(r2^2-r1^2)		surfarea2/vol2
lam	1600		lambda
saiu	2*pi*r1*Lu		inside upper surface area
sail	2*pi*r1*Ll		inside lower surface area
saou	2*pi*r2*Lu		outside upper surface area
saol	2*pi*r2*Ll		outside lower surface area
Vuj	pi*(r2^2-r1^2)*Lu		
Vlj	pi*(r2^2-r1^2)*Ll		
aiu	saiu/Vuj		
aou	saou/Vuj		
ail	sail/Vlj		
aol	saol/Vlj		

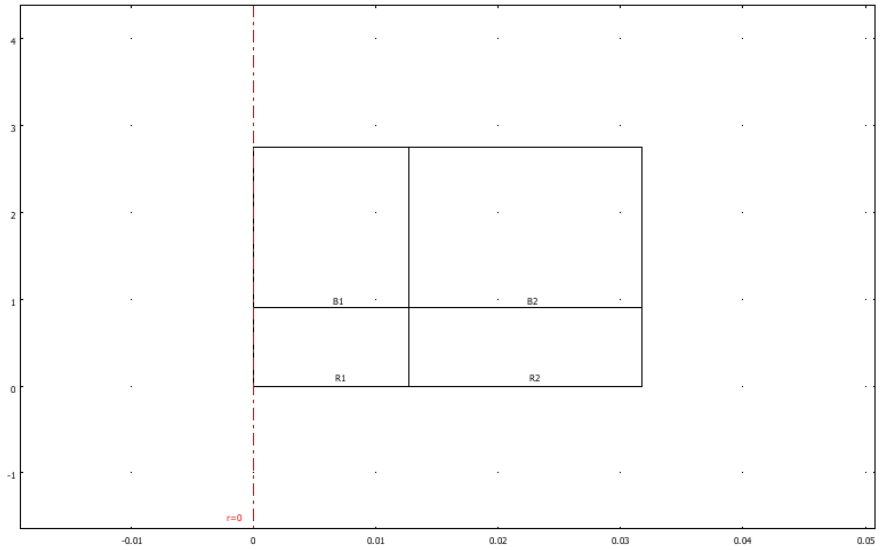
4. Global Expressions

Name	Expression	Unit	Description
Ql	Ul*(Ts-Tp)		
Qu	Uu*(Ts-Tb)		
Qe	Ue*(Ta-Ts)		
erate	Uu*a*(Ts-Tb)/(lam*18)		
Wg	Cg*92/(Cg*92+Cw*18)	1	
Qrate	Uu*(Ts-Tb)*a		

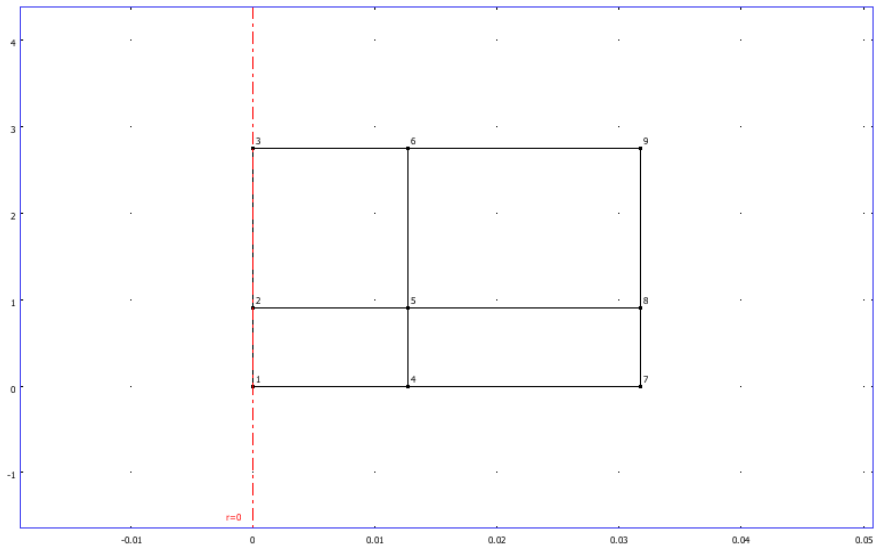
5. Geometry

Number of geometries: 1

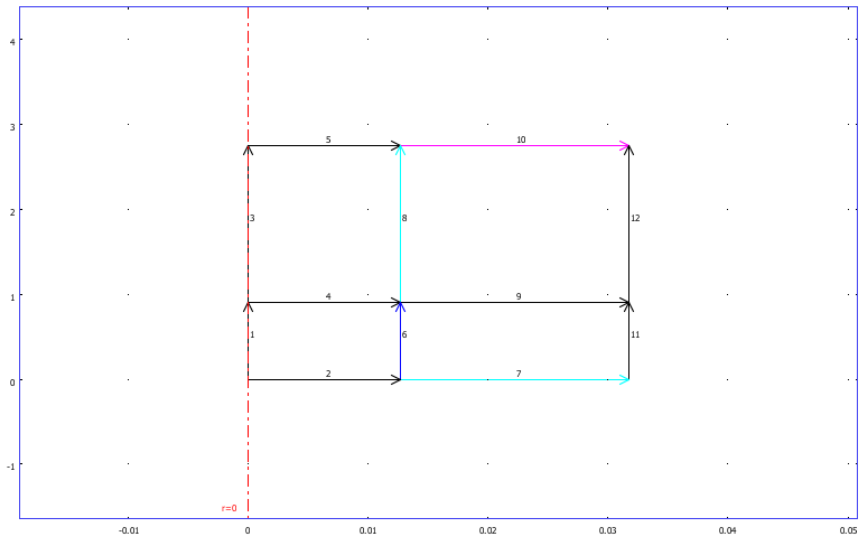
5.1. Geom1



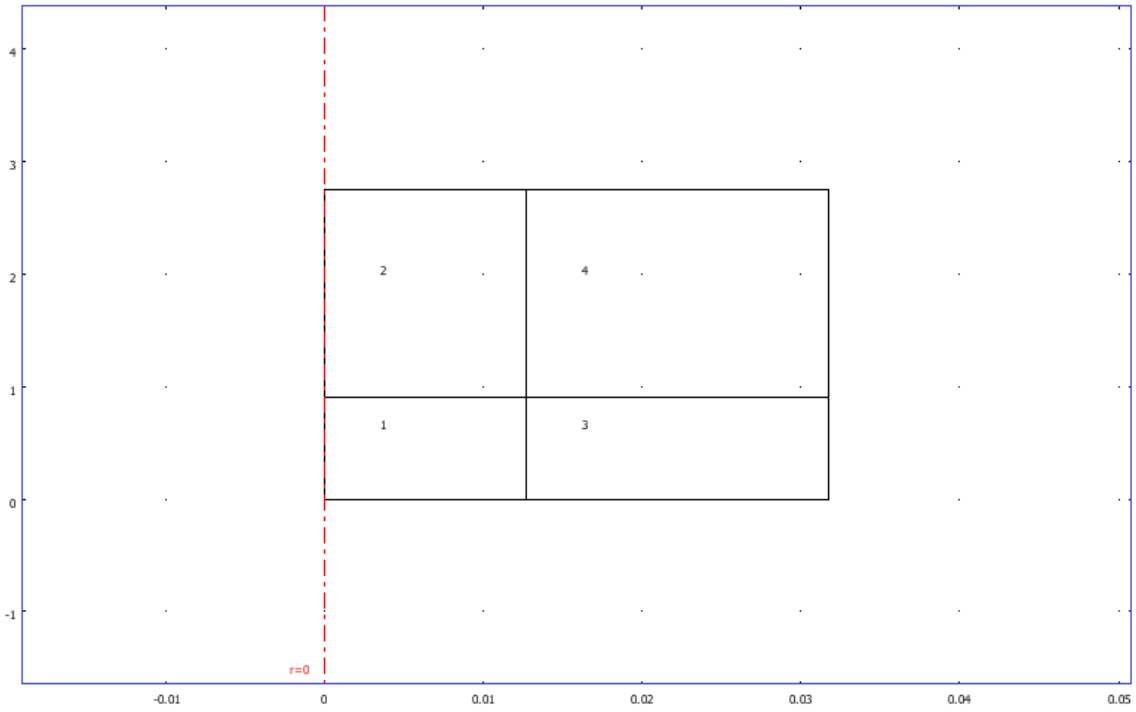
5.1.1. Point mode



5.1.2. Boundary mode



5.1.3. Subdomain mode



6. Geom1

Space dimensions: Axial symmetry (2D)

Independent variables: r, phi, z

6.1. Expressions

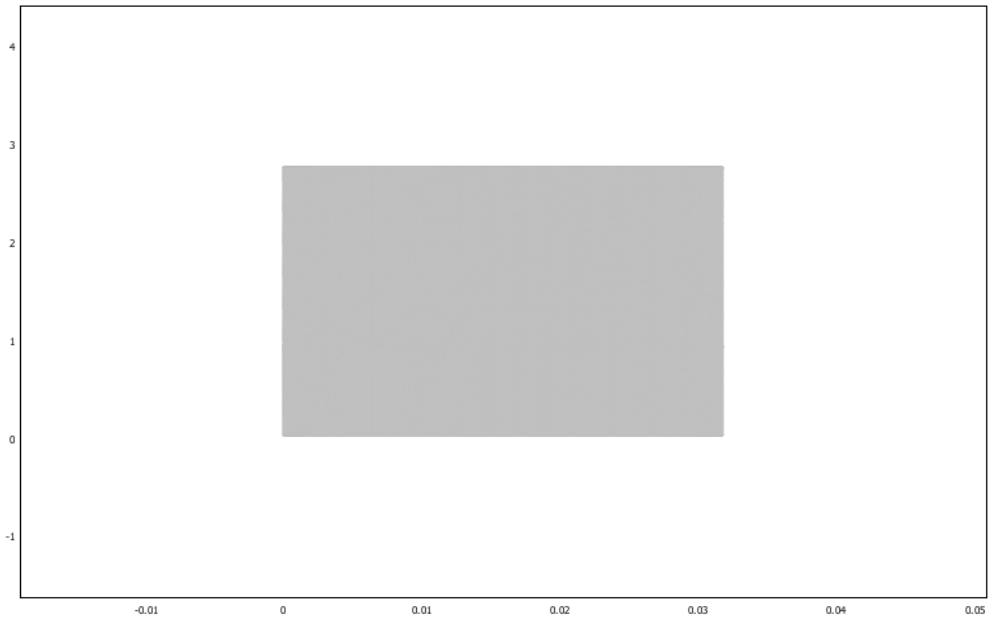
6.1.1. Subdomain Expressions

Subdomain		1	3
Tp		T	Ti

6.2. Mesh

6.2.1. Mesh Statistics

Number of degrees of freedom	68993
Number of mesh points	8121
Number of elements	15104
Triangular	15104
Quadrilateral	0
Number of boundary elements	1788
Number of vertex elements	9
Minimum element quality	0.704
Element area ratio	0.065



6.3. Application Mode: Convection and Diffusion (chcd)

Application mode type: Convection and Diffusion (Chemical Engineering Module)

Application mode name: chcd

6.3.1. Application Mode Properties

Property	Value
Default element type	Lagrange - Quadratic
Analysis type	Stationary
Equation form	Non-conservative
Equilibrium assumption	Off
Frame	Frame (ref)
Weak constraints	Off
Constraint type	Ideal

6.3.2. Variables

Dependent variables: Cw, Cg, Cwv

Shape functions: shlag(2,'Cw'), shlag(2,'Cg'), shlag(2,'Cwv')

Interior boundaries not active

6.3.3. Boundary Settings

Boundary		1, 3	2	5
Type		Axial symmetry	Concentration	Convective flux
Concentration (c0)	mol/m^3	{0;0;0}	**{Cw0;Cg0;Cwv0}**	{0;0;0}

Boundary		6, 8
Type		Insulation/Symmetry
Concentration (c0)	mol/m^3	{0;0;0}

6.3.4. Subdomain Settings

Subdomain		1	2
Diffusion coefficient (D)	m^2/s	**{D;D;Dg}**	**{D;D;Dg}**
Reaction rate (R)	$mol/(m^3 \cdot s)$	{0;0;0}	**{-erate;0;erate}**
z-velocity (v)	m/s	**{vin;vin;vin}**	**{vin;vin;vin}**

Subdomain initial value		1	2
Concentration, Cw (Cw)	mol/m^3	Cw0	Cw0
Concentration, Cg (Cg)	mol/m^3	Cg0	Cg0
Concentration, Cwv (Cwv)	mol/m^3	Cwv0	Cwv0

6.4. Application Mode: Convection and Conduction (chcc)

Application mode type: Convection and Conduction (Chemical Engineering Module)

Application mode name: chcc

6.4.1. Scalar Variables

Name	Variable	Value	Unit	Description
Rg	Rg_chcc	8.31451	J/(mol*K)	Universal gas constant

6.4.2. Application Mode Properties

Property	**Value**
Default element type	Lagrange - Quadratic
Analysis type	Stationary
Species diffusion	Inactive
Turbulence model	None
Predefined multiphysics application	Off
Frame	Frame (ref)
Weak constraints	Off
Constraint type	Ideal

6.4.3. Variables

Dependent variables: T

Shape functions: shlag(2,'T')

Interior boundaries active

6.4.4. Boundary Settings

Boundary		1, 3	2	4, 9-10
Type		Axial symmetry	Temperature	Continuity
Inward heat flux (q0)	W/m^2	0	0	0
Temperature (T0)	K	273.15	**Tf**	273.15

Boundary		5	8	7
Type		Convective flux	Heat flux	Temperature
Inward heat flux (q0)	W/m^2	0	**Qu**	0
Temperature (T0)	K	273.15	273.15	**Ts**

Boundary		11-12	6
Type		Continuity	Heat flux
Inward heat flux (q0)	W/m^2	**-Ue*(Ts-Ta)**	**Ql**

Temperature (T0)	K	273.15	273.15

6.4.5. Subdomain Settings

Subdomain		1	2
Thermal conductivity (k)	W/(m·K)	**K**	**K**
Thermal conductivity (ktensor)	W/(m·K)	**{6000,0;0,.006}**	**{6000,0;0,.006}**
ktype		**aniso**	**aniso**
Density (rho)	kg/m^3	**rho**	**rho**
Heat capacity at constant pressure (C)	J/(kg·K)	**Cp**	**Cp**
Heat source (Q)	W/m^3	0	**-Qrate**
z-velocity (v)	m/s	**vin**	**vin**

Subdomain initial value		1	2
Temperature (T)	K	Tf	Tb

6.5. Application Mode: Convection and Conduction (chcc2)

Application mode type: Convection and Conduction (Chemical Engineering Module)

Application mode name: chcc2

6.5.1. Scalar Variables

Name	Variable	Value	Unit	Description
Rg	Rg_chcc2	8.31451	J/(mol*K)	Universal gas constant

6.5.2. Application Mode Properties

Property	**Value**
Default element type	Lagrange - Quadratic
Analysis type	Stationary
Species diffusion	Inactive
Turbulence model	None
Predefined multiphysics application	Off
Frame	Frame (ref)
Weak constraints	Off
Constraint type	Ideal

6.5.3. Variables

Dependent variables: T2

Shape functions: shlag(2,'T2')

Interior boundaries not active

6.5.4. Boundary Settings

Boundary		6	7	8
Type		Heat flux	Temperature	Heat flux
Inward heat flux (q0)	W/m^2	**-Ql**	0	**-Qu**
Temperature (T0)	K	273.15	**Ts**	273.15

Boundary		10	11-12
Type		Convective flux	Heat flux
Inward heat flux (q0)	W/m^2	0	**Qe**
Temperature (T0)	K	273.15	273.15

6.5.5. Subdomain Settings

Subdomain		3	4
Thermal conductivity (k)	W/(m·K)	**25000**	**25000**
Density (rho)	kg/m^3	**0.6**	**0.6**
Heat capacity at constant pressure (C)	J/(kg·K)	**2058**	**2058**
Heat source (Q)	W/m^3	**Ql*ail-Qe*aol**	**Qu*aiu-Qe*aou**
z-velocity (v)	m/s	**vst**	**vst**

Subdomain initial value		3	4
Temperature (T2)	K	Ts	Ts

7. Extrusion Coupling Variables

7.1. Geom1

7.1.1. Source Subdomain: 1

Name	**Value**
Expression	T
Transformation type	Linear
Destination Subdomain	3 (Geom1)
Source vertices	1, 2
Destination vertices	4, 5
Name	Ti

8. Solver Settings

Solve using a script: off

Analysis type	Stationary
Auto select solver	On
Solver	Stationary
Solution form	Automatic
Symmetric	auto
Adaptive mesh refinement	Off
Optimization/Sensitivity	Off
Plot while solving	Off

8.1. Direct (UMFPACK)

Solver type: Linear system solver

Parameter	Value
Pivot threshold	0.1
Memory allocation factor	0.7

8.2. Stationary

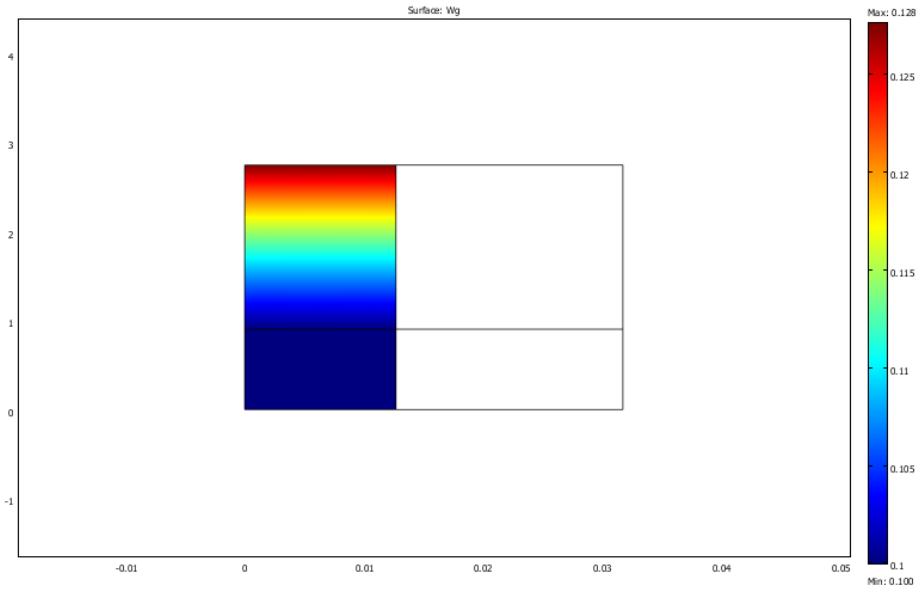
Parameter	Value
Linearity	Automatic
Relative tolerance	1.0E-6
Maximum number of iterations	25
Manual tuning of damping parameters	Off
Highly nonlinear problem	Off
Initial damping factor	1.0
Minimum damping factor	1.0E-4
Restriction for step size update	10.0

8.3. Advanced

Parameter	Value
Constraint handling method	Elimination
Null-space function	Automatic

Automatic assembly block size	On
Assembly block size	1000
Use Hermitian transpose of constraint matrix and in symmetry detection	Off
Use complex functions with real input	Off
Stop if error due to undefined operation	On
Store solution on file	Off
Type of scaling	Automatic
Manual scaling	
Row equilibration	On
Manual control of reassembly	Off
Load constant	On
Constraint constant	On
Mass constant	On
Damping (mass) constant	On
Jacobian constant	On
Constraint Jacobian constant	On

9. Postprocessing



Appendix C- Extra Graphs and Tables

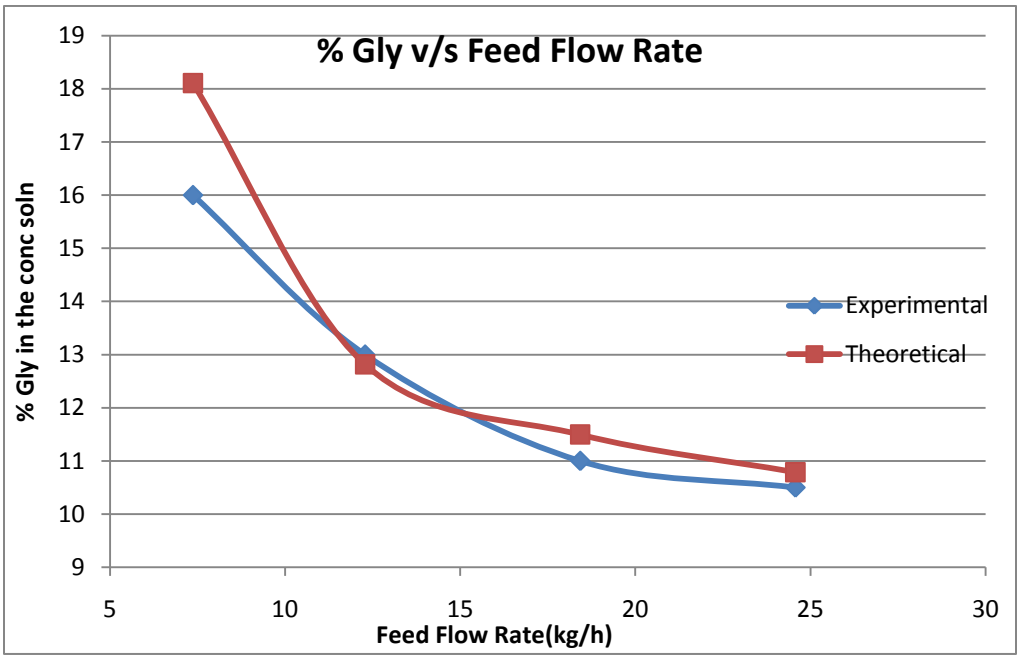


Figure 21: Percentage glycerol v/s feed flow rate at 5 psig

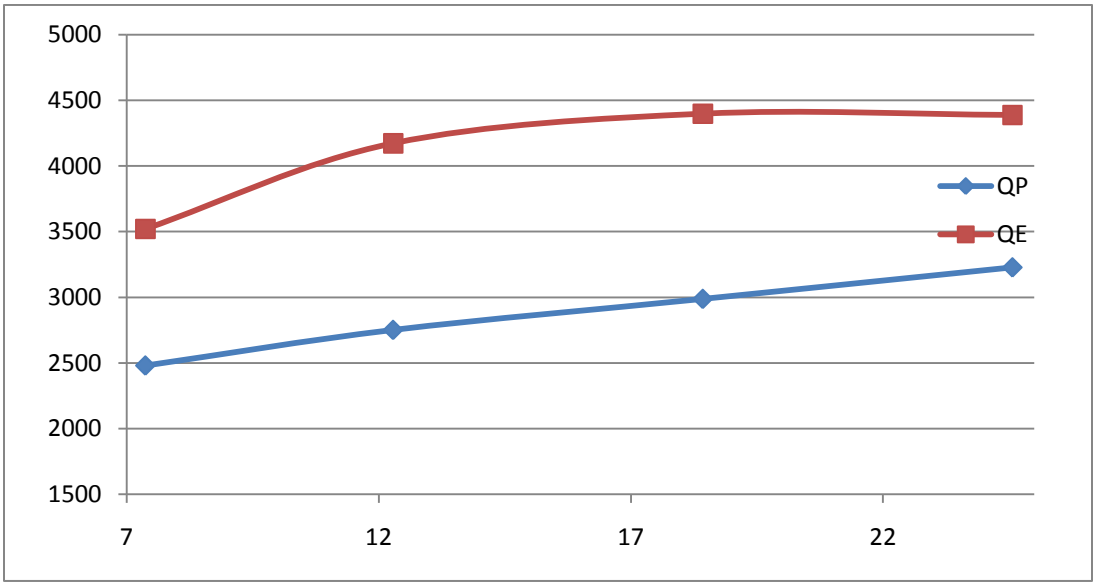


Figure 22: Comparison of QP and QE with varying flow rate at 10 psig

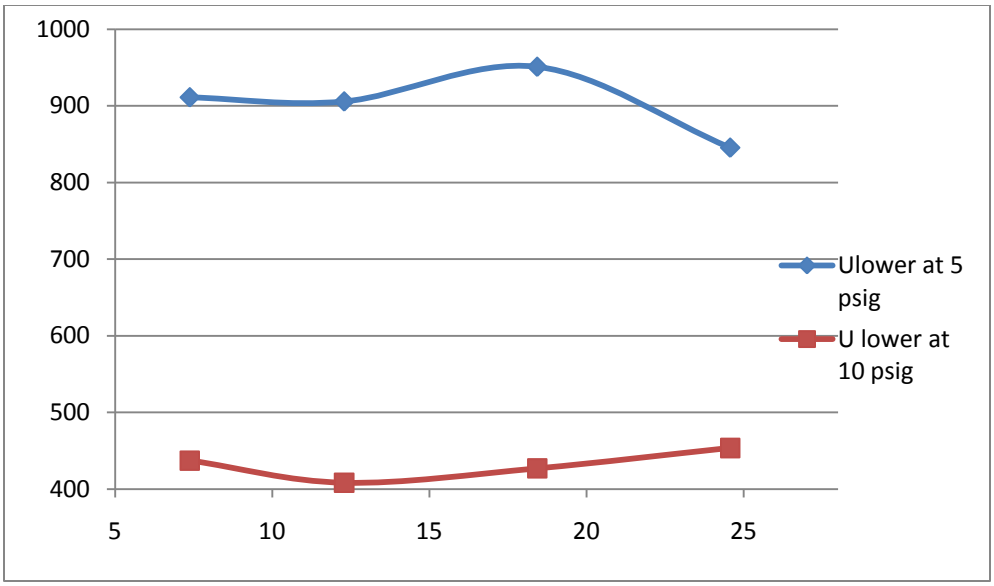
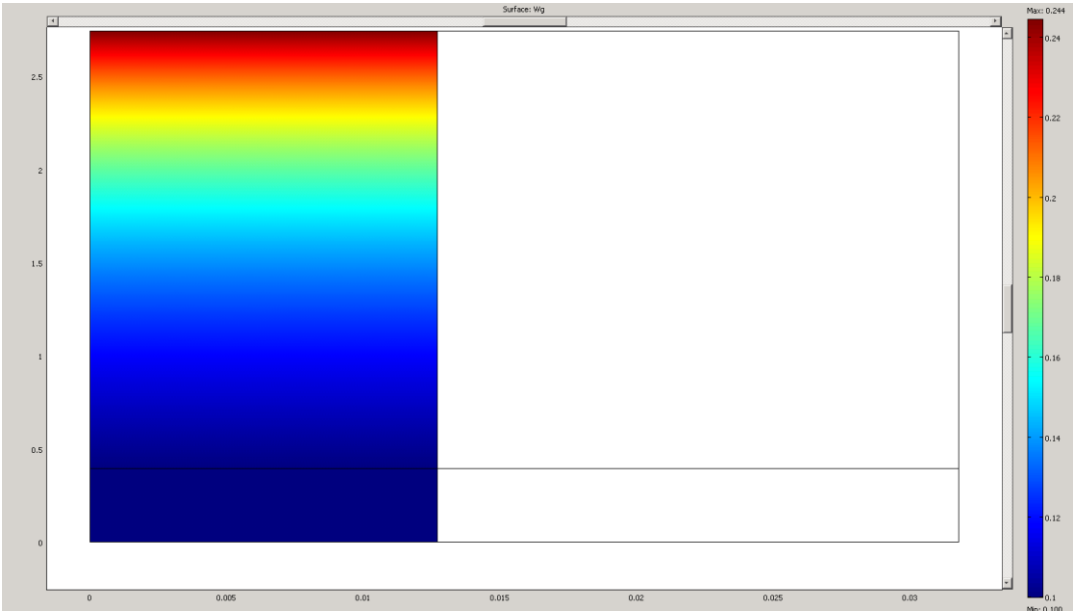
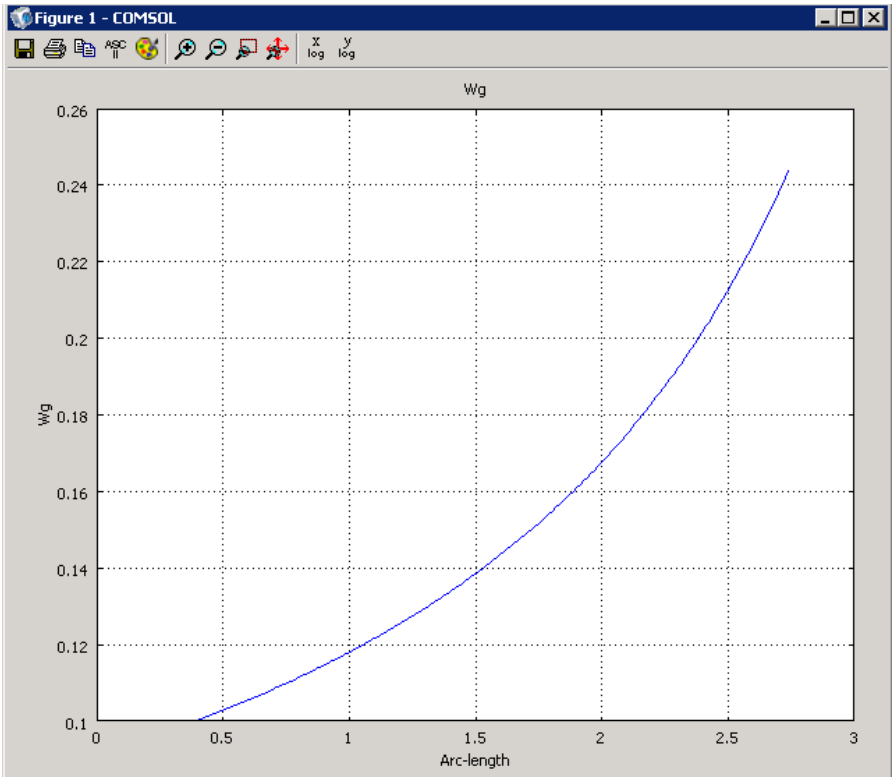


Figure 23: Comparison of Ulower at different steam pressures

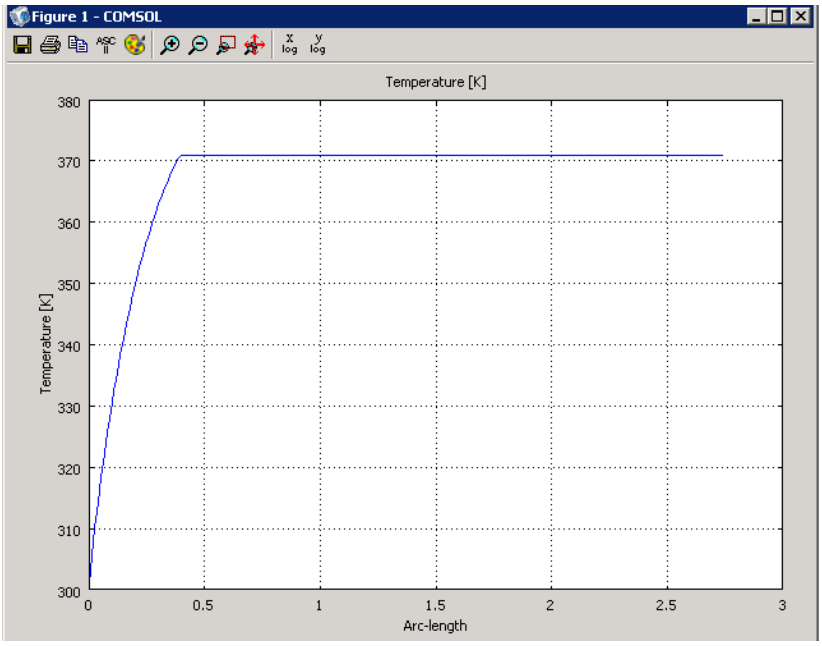
Wg at 120



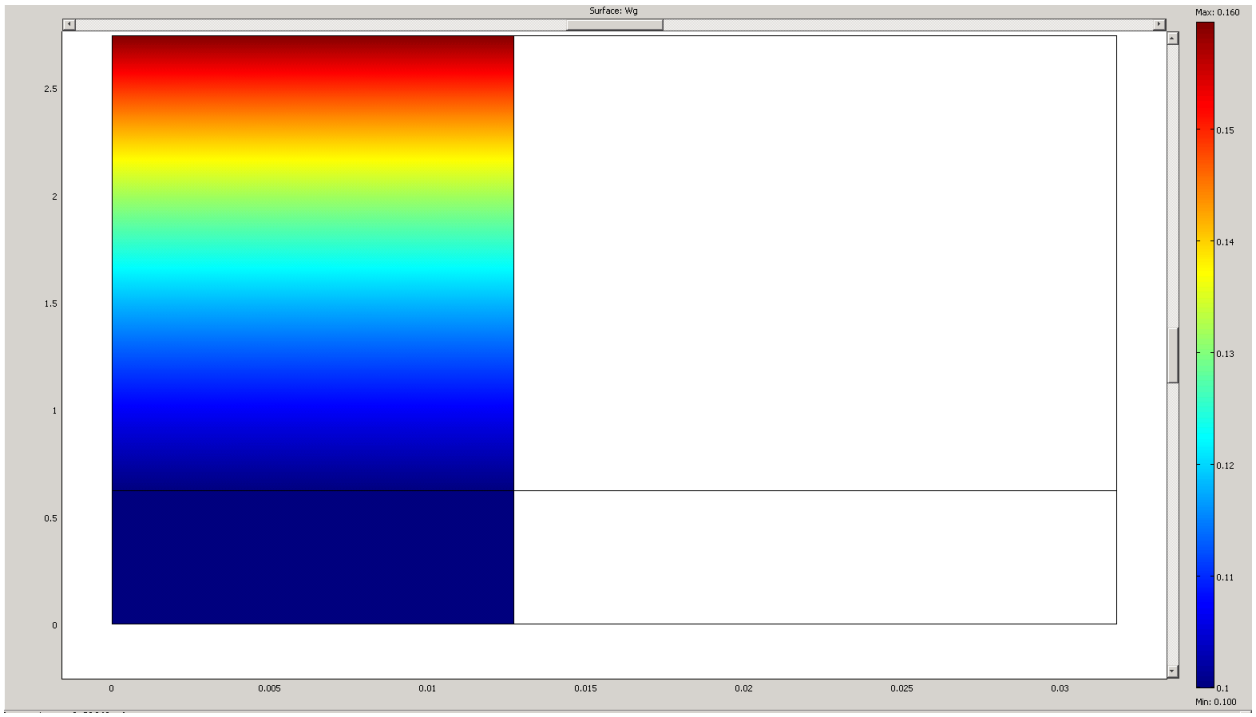
Wg at 120



T at 120



Wg at 200


Surface: Wg
Max: 0.160
Min: 0.100
2.5
2
1.5
1
0.5
0
0
0.005
0.01
0.015
0.02
0.025
0.03
0.15
0.14
0.13
0.12
0.11
0.1

Wg at 200

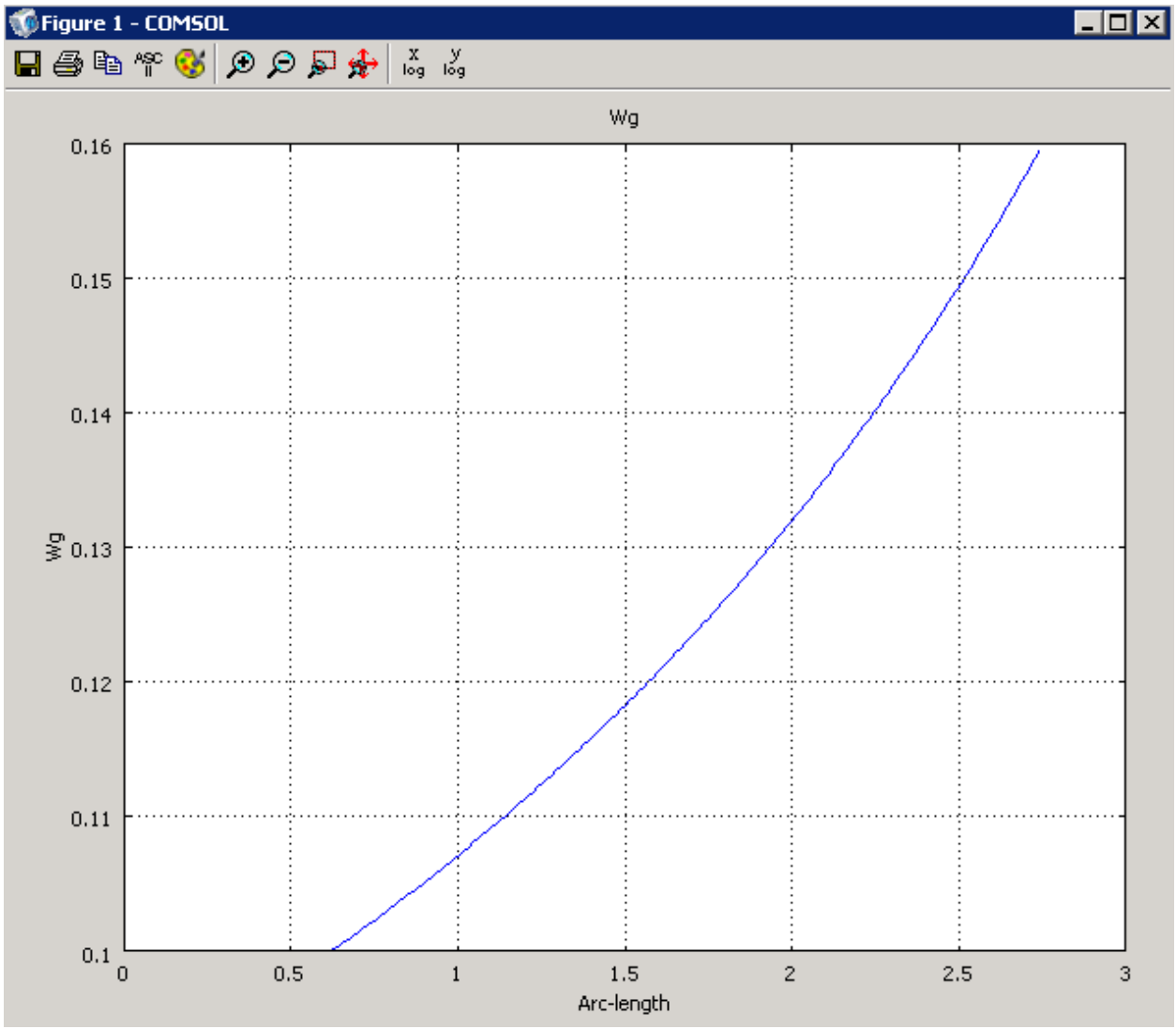


Figure 1 - COMSOL
Wg
0.16
0.15
0.14
0.13
0.12
0.11
0.1
Wg
0
0.5
1
1.5
2
2.5
3
Arc-length

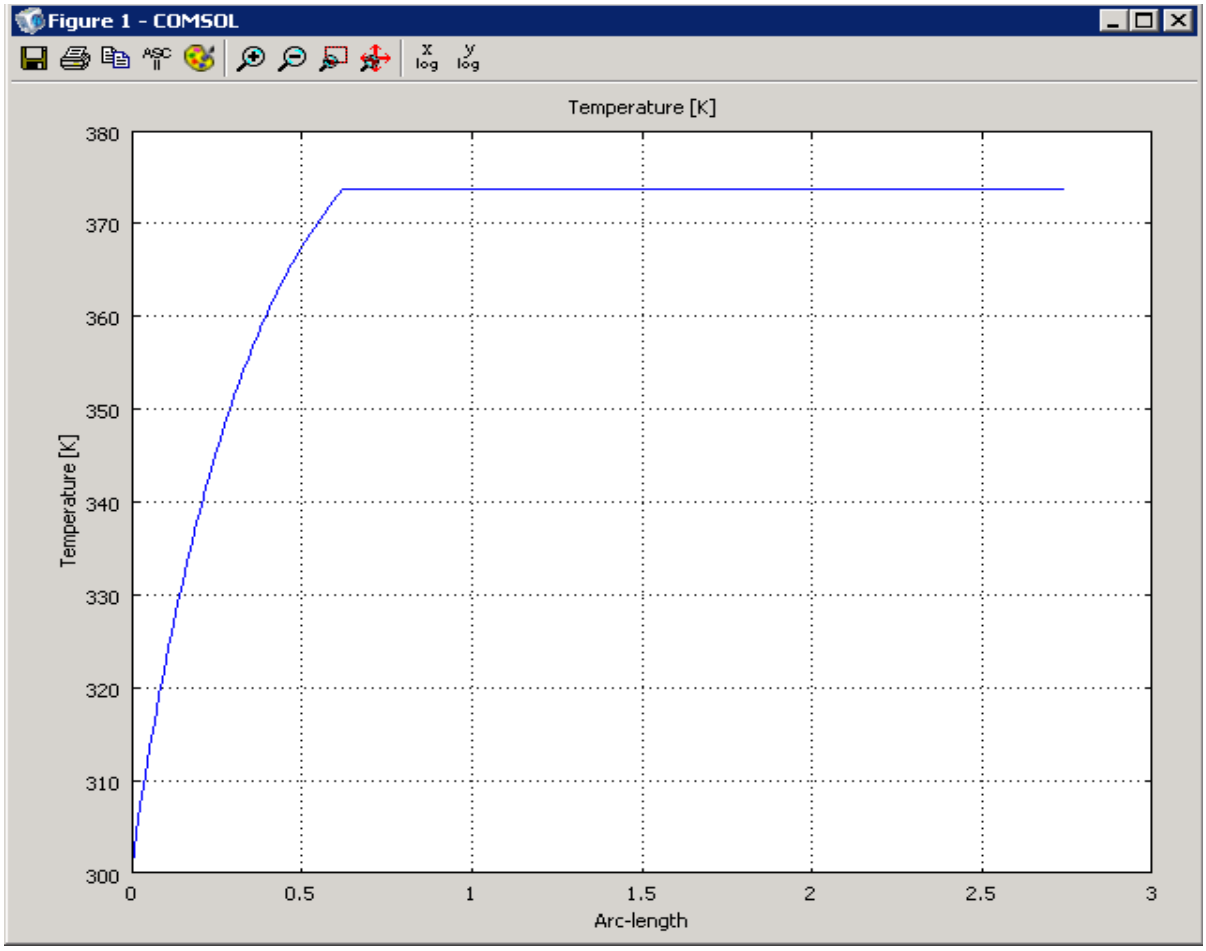
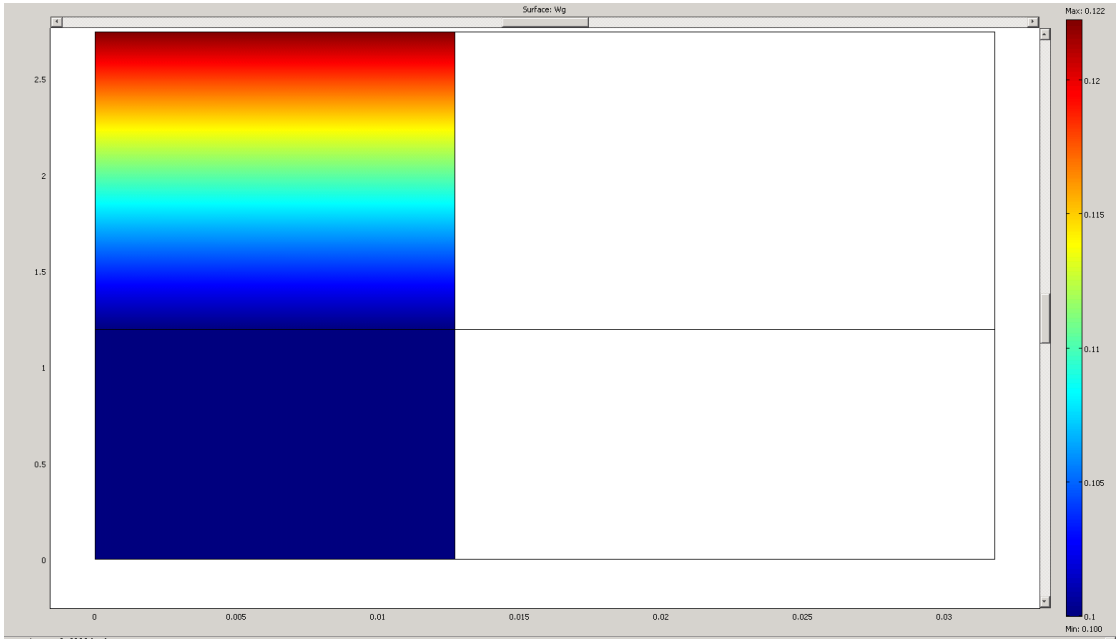
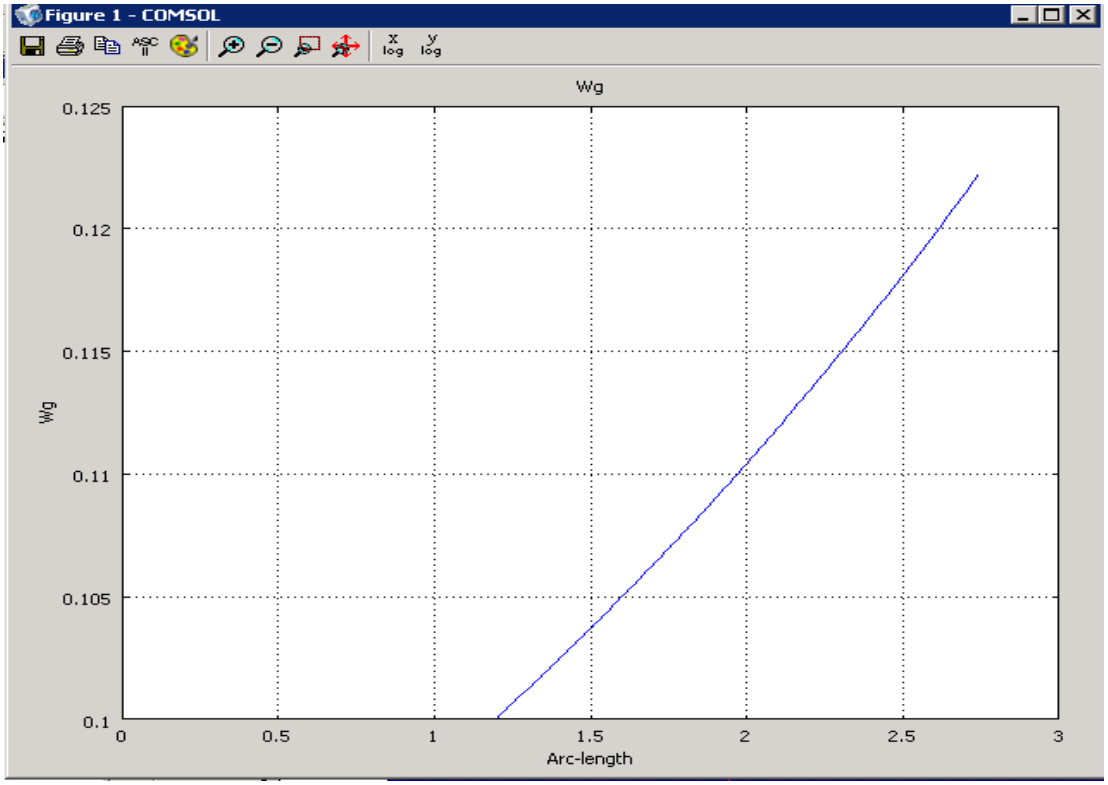


Figure 1 - COMSOL
Temperature [K]
380
370
360
350
340
330
320
310
300
Temperature [K]
0
0.5
1
1.5
2
2.5
3
Arc-length

Wg at 400


Surface: Wg
Max: 0.122
0.12
0.115
0.11
0.105
0.1
Min: 0.100
2.5
2
1.5
1
0.5
0
0
0.005
0.01
0.015
0.02
0.025
0.03

Wg at 400



T at 400

Appendix D- Raw Data and Spreadsheet containing calculations

Column Data

Run(flow rate	Room Temp	Gly feed Flow	Gly Feed T	Cond Flow	Conc Gly Flow	Conc Gly SG	Wt %	BP Height	Temp @ top of c	Steam P	Solun from condenser
	C	ml/min	(F)	g/min	g/min	g/cm^3			(F)		g/min
1		120	82	125	95	1.036	16	16.7	212	10	55
2		200	82.5	135	180	1.027	13	28	212	10	45
3		300	82	145	330	1.024	11	40	212	10	40
4		400	82	155	400	1.021	10.5	72	212	10	30

Measured Mass Flow Rates

Run(flow rate	Room Temp	Wt of the Cond-	Wt of the red f	Wt of cond	Wt of Gly+Buc	Wt of the wh	Wt of gly	Wt of soln	Wt of bu	Wt of soln
	C		(g)			(g)				
1		740	490	250	1140	950	190	1040	930	110
2		760	490	270	1310	950	360	1020	930	90
3		780	490	290	1610	950	660	1010	930	80
4		800	490	310	1750	950	800	990	930	60

Column Data

Run(flow rat	Gly feed F	Gly Feed T	Cond Flow	Conc Gly Fl	Conc Gly S	Wt %	BP Height	Temp @ t	Steam P	Solun from condenser
	ml/min	(F)	g/min	g/min	g/cm^3			(F)		g/min
1	120	83	130	80	1.039	17	15.5	212	15	70
2	200	83.5	150	155	1.031	15	24.5	212	15	65
3	300	84	160	285	1.027	13	35.5	212	15	55
4	400	84	165	400	1.025	12	47	212	15	45

Measured Mass Flow Rates

Run(flow rat	Wt of the	Wt of the re	Wt of cond	Wt of Gly+	Wt of the	Wt of gly	Wt of soln+bu	Wt of bu	Wt of soln
		(g)			(g)				
1	750	490	260	1190	1030	160	1070	930	140
2	790	490	300	1340	1030	310	1060	930	130
3	810	490	320	1600	1030	570	1040	930	110
4	820	490	330	1830	1030	800	1020	930	90

DAY 1

Calculations using density

120							
200							
300							
400							Density
Density of g	%	%	Den due t	Density of	% of Wtar	Den W	Total
1.261	10	0.1	0.1261	0.9974	0.9	0.89766	1.02376
1.261	10	0.1	0.1261	0.9974	0.9	0.89766	1.02376
1.261	10	0.1	0.1261	0.9974	0.9	0.89766	1.02376
1.261	10	0.1	0.1261	0.9974	0.9	0.89766	1.02376

Flow Rate IN

ml/min	density	g/min	kg/h	% Gly	%	Wt gly	Wt of Water
120	1.02376	122.8512	7.371072	10	0.1	0.737107	6.633965
200	1.02376	204.752	12.28512	10	0.1	1.228512	11.05661
300	1.02376	307.128	18.42768	10	0.1	1.842768	16.58491
400	1.02376	409.504	24.57024	10	0.1	2.457024	22.11322

Calculations for condenser soln(mainly water)

Flow Rate out	g/min	kg/h	Gly	Water	
	55	3.3	0	3.214264	3.214264
	45	2.7	0	2.621702	2.621702
	40	2.4	0	2.320277	2.320277
	30	1.8	0	1.733851	1.733851

(Calculations for Conc Gly)

Flow Rate Out	g/min	kg/h	Gly %	%	Wt of Glu	Wt of Water
		4.071072	16	0.16	0.651372	3.4197
		9.58512	12	0.12	1.150214	8.434906
		16.02768	11	0.11	1.763045	14.26464
		22.77024	10.5	0.105	2.390875	20.37936

[redacted]

Theoretical Calculations

mass flow	%gly in fe	mass flow	% gly in li	% Gly
7.371072	0.1	4.071072	0.18106	18.10597
12.28512	0.1	9.58512	0.128169	12.81687
18.42768	0.1	16.02768	0.114974	11.49741
24.57024	0.1	22.77024	0.107905	10.79051

Mass Balance for steam

Out		In(assumed)
125	7.5	7.5
135	8.1	8.1
150	9	9
155	9.3	9.3

Calculations using density

120							
200							
300							
400							Density
Density of g	%		%	Den due t	Density of	% of Wtar	Den W	Total
1.261	10		0.1	0.1261	0.9974	0.9	0.89766	1.02376
1.261	10		0.1	0.1261	0.9974	0.9	0.89766	1.02376
1.261	10		0.1	0.1261	0.9974	0.9	0.89766	1.02376
1.261	10		0.1	0.1261	0.9974	0.9	0.89766	1.02376

Flow Rate IN

(kg/h)	density	g/min	kg/h	% Gly	%	Wt gly	Wt of Water
120	1.02376	122.8512	7.371072	10	0.1	0.737107	6.633965
200	1.02376	204.752	12.28512	10	0.1	1.228512	11.05661
300	1.02376	307.128	18.42768	10	0.1	1.842768	16.58491
400	1.02376	409.504	24.57024	10	0.1	2.457024	22.11322

Calculations for condenser soln(mainly water)

Flow Rate out	g/min	kg/h	Gly	Water	%gly	% water
	70	4.2	0	4.001975		
	65	3.9	0	3.845405		
	55	3.3	0	3.272554		
	45	2.7	0	2.648702		

(Calculations for Conc Gly)

Flow Rate Out	g/min	kg/h	Gly %	%	Wt of Glu	Wt of Water
		3.171072	17	0.17	0.539082	2.63199
		8.38512	14	0.14	1.173917	7.211203
		15.12768	12	0.12	1.815322	13.31236
		21.87024	11	0.11	2.405726	19.46451

[redacted]

Theoretical Calculations

mass flow	%gly in fe	mass flow	% gly in li	% Gly
7.371072	0.1	3.171072	0.232447	23.24473
12.28512	0.1	8.38512	0.146511	14.6511
18.42768	0.1	15.12768	0.121814	12.502
24.57024	0.1	21.87024	0.112346	11.23455

Mass Balance for steam

Out		In(assumed)	
130	7.8	7.8	
150	9	9	
160	9.6	9.6	
165	9.9	9.9	

Day 1@ 5 psig

Heat Gained by the Process

kg/h	kj/kg	J/kg	kg/h	%	J/mol.K	g	J/g.k	J/kg.k	%	J/kg.k		F	K	F	K	J/h	J/s	kW
M (cond soln	Lambda		M(vap)	%liq	Cp(gly)	MM of gly	Cp(gly)	Cp(gly)	% Water	Cp(water)	Cp aveg	boiling	Tout	T(in)	T(in)	Qp	Qp	Qp
7.371072	1600	1600000	3.3	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4180	373.15	374.15	82	300.93	7535987	2093.33	2.09333
12.28512	1600	1600000	2.7	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4180	373.15	374.15	82.5	301.21	8014249	2226.18	2.22618
18.42768	1600	1600000	2.4	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4180	373.15	374.15	82	300.93	9402941	2611.928	2.611928
24.57024	1600	1600000	1.8	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4008	373.15	374.15	82	300.93	9992047	2775.569	2.775569

Quality of the Steam

Hcond	Hsuper	Hcond	Hsat	Quality
180.16	1157.42	180.16	1150.5	1
198.32	1157.42	198.32	1157.1	1

Heat lost by steam

kg/h	kg/sec		X	Watts	Watts	watts	
Mass flow of steam		Lambda	Quality	Qp	QS	QE(heat lc	QE
7.5	0.002083	2234000	1	2093.33	4654.166667	2560.837	2.560837
8.1	0.00225	2234000	1	2226.18	5026.5	2800.32	2.80032
9	0.0025	2234000	1	2611.928	5585	2973.072	2.973072

Day 2@ 10 psig

Heat Gained by the Process												Boiling	Outlet					
kg/h	kj/kg	J/kg	kg/h	%	J/mol.K	g	J/g.k	J/kg.k	%	J/kg.k		F	K	F	K	J/h	J/s	kW
M (cond soln	Lambda		M(vap)	%liq	Cp(gly)	MM of gly	Cp(gly)	Cp(gly)	% Water	Cp(water)	Cpaveg	Tboiling)	Tb	T(feed)	T(feed)	Qp	Qp	Qp
7.371072	1600	1600000	4.2	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4180	373.15	376.15	83	301.48	9020663	2505.74	2.50574
12.28512	1600	1600000	3.9	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4180	373.15	376.15	83.5	301.76	10060061	2794.461	2.794461
18.42768	1600	1600000	3.3	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4180	373.15	376.15	84	302.04	10988523	3052.368	3.052368
24.57024	1600	1600000	2.7	0.1	221.9	92.09382	2.409499	2409.499	0.9	4186	4008	373.15	376.15	84	302.04	11618169	3227.269	3.227269

Quality of the Steam

Hcond	Hsuper	Hcond	Hsat	Quality
180.16	1160.44	180.16	1150.5	1
208.44	1160.74	208.44	1160.7	1

Heat lost by steam

kg/h	kg/sec		X	Watts	Watts	watts	kwatts
Mass flow of steam		Lambda	Quality	Qp	QS	Heat lost	QE
7.8	0.002167	2216000	1	2505.74	4801.333333	2295.593	2.295593
9	0.0025	2216000	1	2794.461	5540	2745.539	2.745539
9.6	0.002667	2216000	1	3052.368	5909.333333	2856.966	2.856966
9.9	0.00275	2216000	1	3227.269	6094	2866.731	2.866731

Day 1@ 5 psig

Calculations for Overall Heat Transfer Coefficient					Air		Feed		Steam		Outlet (tube)		Boiling				
m	m	m	Inner (m2	Outer	F	Kelvin	F		F		Toutlet		Tboiling	QE	Ts-Ta	Outer SA	
Pipe Dia(in)	Pipe Dia(out)	Length	Area(oute	Area	Tair	Tair	Tinlet	Kelvin	Ts(F)	Kelvin	F	Kelvin	Kelvin				U
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	227	381.483	212	374.15	373.15	2560.837	83.33333	0.546605	56.21981
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82.5	300.927778	227	381.483	212	374.15	373.15	2800.32	83.33333	0.546605	61.47734
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	227	381.483	212	374.15	373.15	2973.072	83.33333	0.546605	65.26989
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	227	381.483	212	374.15	373.15	2995.598	83.33333	0.546605	65.76442

Day 2@ 10 psig

Calculations for Overall Heat Transfer Coefficient					Air		Feed		Steam		Outlet (tube)		Boiling				
m	m	m	Inner (m2	Outer	F	Kelvin	F		F		Toutlet		Tboiling	QE	Ts-Ta	Outer SA	
Pipe Dia(in)	Pipe Dia(out)	Length	Area(oute	Area	Tair	Tair	Tinlet	Kelvin	Ts(F)	Kelvin	F	Kelvin	Kelvin				U
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	239	389.15	212	376.15	373.15	2295.593	91	0.546605	46.15086
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	239	389.15	212	376.15	373.15	2745.539	91	0.546605	55.19661
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	239	389.15	212	376.15	373.15	2856.966	91	0.546605	57.43675
0.0254	0.0635	2.74	0.000507	0.003167	77	298.15	82	300.927778	239	389.15	212	376.15	373.15	2866.731	91	0.546605	57.63307

U lower & upper calculations													
5 Psig	Height	Inner	Outer							Inner	Outer		
inches	m	AREA(in^2)	Area (m^2	m	Cp(feed)	Tb-TF	q lower	Qlower	LMTD lo	U(lower)			
16.7	0.42418	0.033848	0.08462	7.371072	4180	72.22222	2225244.67	618.123519	76.8824	237.528	95.01128		
28	0.7112	0.0567512	0.141878	12.28512	4180	72.22222	3708741.11	1030.20586	76.8824	236.114	94.44574		
40	1.016	0.0810731	0.202683	18.42768	4008	72.22222	5334198.94	1481.72193	76.8824	237.719	95.08743	300	0.207605
60	1.397	0.1114756	0.278689	24.57024	4180	72.22222	7417482.23	2060.41173	76.8824	240.407	96.16293		
Vapor		Inner	Outer						Inner	Outer			
inches		Area	Area (m^2	m	Cp(feed)	q upper	Qupper	LMTD lower	U(upper)				
91.3	2.3495	0.1874816	0.468704	3.3	1600000	5280000	1466.66667	8.333333	938.759	375.504			
80	2.0209	0.1612605	0.403151	2.7	1600000	4320000	1200	8.333333	892.965	357.186			
68	1.8415	0.146945	0.367363	2.4	1600000	3840000	1066.66667	8.333333	871.074	348.43		900	0.03214
48	1.346	0.1074059	0.268515	1.8	1600000	2880000	800	8.333333	893.805	357.522			
U lower & upper calculatic		Inner	Outer										
10psig		Surface Area								Inner			
inches		AREA(in^2)	Area (m^2	m	Cp(feed)	Tb-TF	q lower	Qlower	LMTD lo	U(lower)	Outer		
15.5	0.397	0.0316792	0.079198	7.371072	4180	75.22222	2317677.91	643.79942	81.4334	249.56	99.82385		
24.5	0.6223	0.0496573	0.124143	12.28512	4180	75.22222	3862796.52	1072.99903	81.4334	265.347	106.1387		
35.5	0.9017	0.0719524	0.179881	18.42768	4180	75.22222	5794194.78	1609.49855	81.4334	274.69	109.876	400	0.313275
47	1.1938	0.0952609	0.238152	24.57024	4180	75.22222	7725593.03	2145.99807	81.4334	276.638	110.6552		
									Inner	Outer			
inches		AREA(in^2)	Area (m^2	m	Cp(feed)	q upper	Qupper	LMTD lower	U(upper)		U		
92.5	2.3495	0.1874816	0.468704	4.2	1600000	6720000	1866.66667	16	622.283	248.913			
83.5	2.1209	0.1692402	0.4231	3.9	1600000	6240000	1733.33333	16	640.116	256.046			
72.5	1.8415	0.146945	0.367363	3.3	1600000	5280000	1466.66667	16	623.816	249.526		800	0.22023
61	1.5494	0.1236365	0.309091	2.7	1600000	4320000	1200	16	606.617	242.647			

Day 1		
Calculations for Evaporator Capacity and Economy		
Feed Flow	Evap Capa	Economy
7.371072	3.3	0.44
12.28512	2.7	0.333333
18.42768	2.4	0.266667
24.57024	1.8	0.193548

Day 2		
Calculations for Evaporator Capacity and Economy		
Feed Flow	Evap Capacity=mvap	Economy
7.371072	4.2	0.538462
12.28512	3.9	0.433333
18.42768	3.3	0.34375
24.57024	2.7	0.272727